PB87-166583

EAB(Economic Analysis Branch) Control Cost Manual (Third Edition)

(U.S.) Environmental Protection Agency
Research Triangle Park, NC

Feb 87

U.S. Department of Commerce
National Technical Information Service
NTIS

PB87-166583

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SPRINGFIELD, VA 22161

TECHNICAL REPORT DATA
(Please read Instructions on the reverse before completing)

1. REPORT NO. EPA-450/5-87-001A	2.	3. RECIPIENT'S ACCESSION NO. PB87-166583/AS
4. TITLE AND SUBTITLE EAB Control Cost Manual (Third Edition)		5. REPORT DATE February 1987 6. PERFORMING ORGANIZATION CODE
7. AUTHOR(S) W.M. Vatavuk (principal), R.E. Jenkins, V.S. Katari, W.L. Klotz, J.D. McKenna, R.L. Stallings J.H. Turner, A.S. Viner		8. PERFORMING ORGANIZATION REPORT NO.
9. PERFORMING ORGANIZATION NAME AND ADDRESS Office of Air and Radiation Office of Air Quality Planning and Standards Strategies and Air Standards Division Economic Analysis Branch MD-12 Research Triangle Park, NC 27711		10. PROGRAM ELEMENT NO. 11. CONTRACT/GRANT NO.
12. SPONSORING AGENCY NAME AND ADDRESS Same as box 9.		13. TYPE OF REPORT AND PERIOD COVERED 14. SPONSORING AGENCY CODE

15. SUPPLEMENTARY NOTES
Principal Contact: William M. Vatavuk (Comm'l #: 919-541-5309; FTS # 629-5309)

16. ABSTRACT

This report provides design procedures and current cost data for commonly used "add-on" air pollution control devices and auxiliary equipment. It supercedes two EPA reports, both entitled Capital and Operating Costs of Selected Air Pollution Control Systems: EPA-450/3-76-014 (May 1976) and its revision, EPA-450/5-80-002 (December 1978). This issuance of the Manual contains five sections. Sections 1 ("Introduction") and 2 ("Manual Estimating Methodology") present, respectively, introductory material and a comprehensive, general procedure for developing "study" (± 30% accurate) control cost estimates. Sections 3 to 5 discuss the applications, design, and cost estimating procedures and data for thermal and catalytic incinerators, carbon adsorbers, and fabric filters, respectively. This issuance of the Manual will be followed by others as more sections are completed.

17. KEY WORDS AND DOCUMENT ANALYSIS

a. DESCRIPTORS	b. IDENTIFIERS/OPEN ENDED TERMS	c. COSATI Field/Group
Costs Manual Fabric filters Thermal incinerators Catalytic incinerators Carbon adsorbers	Cost estimation techniques Capital costs Annual costs Control device design and operation Air pollution control systems	

18. DISTRIBUTION STATEMENT Unlimited	19. SECURITY CLASS *(This Report)* Unclassified	21. NO. OF PAGES 184
	20. SECURITY CLASS *(This page)* Unclassified	22. PRICE

EPA Form 2220-1 (Rev. 4-77) PREVIOUS EDITION IS OBSOLETE

NOTE:

- A preliminary draft of this Manual was distributed in November 1986. Since then, extensive revisions have been made to Section 5 ("Fabric Filters") and minor changes to Sections 3 ("Thermal and Catalytic Incinerators") and 4 ("Carbon Adsorbers"). No revisions were made to Sections 1 and 2, however.

- The pages in this Manual have been fastened together. If desired, they may be unfastened, hole-punched, and reassembled in a three-ring binder. This will facilitate the addition of updates when they become available.

OVERVIEW OF THE MANUAL*

	Section	Completion Date	Date To Be Completed
1	INTRODUCTION	1/86	
2	MANUAL ESTIMATING METHODOLOGY	1/86	
3	THERMAL AND CATALYTIC INCINERATORS	7/86	
4	CARBON ADSORBERS	7/86	
5	FABRIC FILTERS	9/86	
6	ELECTROSTATIC PRECIPITATORS		FY 87
7	DUCTWORK		FY 87
8	FANS		FY 87
9	FLARES		FY 87
10	REFRIGERATION UNITS		FY 87
11	STACKS		FY 88
12	PUMPS		FY 88
13	GAS ABSORBERS		FY 88
14	VENTURI SCRUBBERS		FY 88
15	WASTEWATER TREATMENT FACILITIES**		FY 88
16	CAPTURE HOODS		Not Established
17	MECHANICAL COLLECTORS		Not Established
18	PRECOOLERS (SPRAY CHAMBERS AND QUENCHERS)		Not Established
19	SCREW CONVEYORS		Not Established
20	COMPREHENSIVE SAMPLE COST-ESTIMATING PROBLEMS		Not Established
21	ESCALATING COSTS		Not Established
APPENDIX A	COMPOUND INTEREST FACTORS		Not Established
APPENDIX B	EQUIPMENT COST INDICIES		Not Established
APPENDIX C	CONTROL EQUIPMENT VENDORS		Not Established

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*For more information about the Manual, contact Albert Wehe, Robert Pahel-Short, or William Vatavuk (at FTS:629-5610).

**These are facilities that would be required to treat only the wastewater generated by venturi scrubbers, quenchers, spray chambers, and gas absorbers.

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Section 1

INTRODUCTION

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1.1 Role of Cost in the Setting of Regulations

Cost has an important role in setting many state and federal air pollution control regulations. The extent of this role varies with the type of regulation. For some types of regulations, cost is explicitly used in determining their stringency. This use may involve a balancing of costs and environmental impacts, costs and dollar valuation of benefits, or environmental impacts and economic consequences of control costs.

For other types of regulations cost analysis is used to choose among alternative regulations with the same level of stringency. For these regulations, the environmental goal is determined by some set of criteria which do not include costs. However, cost-effectiveness analysis is employed to determine the minimum cost way of achieving the goal.

For some regulations, cost influences enforcement procedures or requirements for demonstration of progress towards compliance with an air quality standard. For example, the size of any monetary penalty assessed for noncompliance as part of an enforcement action needs to be set with awareness of the magnitude of the control costs being postponed by the non-complying facility. For regulations without a fixed compliance schedule, demonstration of reasonable progress towards the goal is sometimes tied to the cost of attaining the goal on different schedules.

Cost is a vital input into two other types of analyses that also sometimes have a role in standard setting. Cost is needed for a benefit-cost analysis that addresses the economic efficiency of alternative regulations.

Cost is also an input into any analysis of the economic impact of each regulatory alternative. An economic impact analysis deals with the consequences of the regulation for small businesses, employment, prices, and industry structure.

1.2 Purpose of Manual

The purpose of this *Manual* is two-fold: (1) to compile up-to-date capital costs, operating and maintenance expenses, and other costs for "add-on" air pollution control systems and (2) to provide a comprehensive, concise, consistent, and easy-to-use procedure for estimating and (where appropriate) escalating these costs. ("Add-on" systems are those installed downstream of an air pollution source to control its emissions.)

The *Manual* estimating procedure rests on the notion of the "factored" or "study" estimate, nominally accurate to within ± 30%. This type of estimate is well suited to estimating control system costs intended for use in regulatory development. Study estimates are sufficiently accurate, yet do not require the detailed, site-specific data inputs needed to make "definitive" or other more accurate types of estimates.

1.3 Organization of the Manual

This *Manual* is a major revision of the 1978 edition of the *EAB Control Cost Manual*,(1) which, in turn, was a revision of the original edition, completed in 1976. This third edition of the *Manual* includes a more thorough discussion of estimating methodology and more detailed design procedures for an enlarged set of equipment types. The appendices have been revised to delete some infrequently used material, and to include certain new and more useful material.

The format of the *Manual* has been changed to one which will permit more flexibility in its updating and expansion. To achieve this flexibility this edition will be issued in self-contained sections. Each section will address a logically separate topic, which can be either of a general nature (e.g., this introduction) or of a more specific, equipment-oriented nature (e.g., fabric filters). The sections which comprise this portion of the Manual are listed in Table 1-1, alongside the sections in the 1978 Manual they will replace.

Two changes in the 1978 Manual are indicated by Table 1-1. First, the numbering scheme is different. In the third edition, each type of equipment, background topic, etc., is given its own number, for ease of identification and to reinforce the intent that each section should "stand alone". Second, the auxiliary equipment items (e.g., ductwork), which were collected into a single chapter in the previous Manuals, are now also stand-alone sections. This was mainly done to eliminate the confusion that arises when classifying auxiliaries like mechanical collectors which can either support a primary control device or be control devices in their own right.

The notion of a stand alone section is also new. Where in the 1976 and 1978 Manuals, the various capital and annual cost factors were collected in an introductory section (old Section 3), now they are dispersed among the sections covering the equipment types. Each of these sections contains a:

- Process description, where the types, uses, and operating modes of the equipment item and (if applicable) its auxiliaries are discussed;
- Design procedure, which enables one to use the parameters of the pollution source (e.g., gas volumetric flowrate) to size the equipment item(s) in question;
- Capital and annual costs for the equipment and suggested factors to use in estimating these costs from equipment design and operational

Table 1-1 Format of the EAB Control Cost Manual (Third Edition)

New Section		Old Section(s) Replaced	
Number	Title	Number(s)	Title
1	"Introduction"	1	"Introduction"
2	"Manual Estimating Methodology"	2,3	"Application to Industry"; "Cost Estimating Procedures"
3	"Thermal and Catalytic Incinerators"	5.4	"Thermal and Catalytic Incinerator Systems"
4	"Carbon Adsorbers"	5.5	"Adsorbers"
5	"Fabric Filters"	5.3	"Fabric Filters"
	(Other sections to be developed)		

(e.g., operating hours) parameters. These costs are presented in both graphical and equational forms wherever possible.

1.4 "Uniqueness" of the Manual

The *Manual* presents a different perspective on estimating air pollution control system costs than other cost-oriented reports, such as:

- *The Cost Digest: Cost Summaries of Selected Environmental Control Technologies*(2)
- *A Standard Procedure for Cost Analysis of Pollution Control Operations*(3)
- *Evaluation of Control Technologies for Hazardous Air Pollutants*(4)

Although these reports (as well as many of the NSPS Background Information Documents) contain costs for add-on control systems, they do not duplicate the *Manual* for one or more of the following reasons: (1) their costs have been based either wholly or partly on data in the previous *Manuals*; (2) they apply to specific source categories only, whereas the *Manual* data may be applied generally; (3) their estimating procedures and costs are of less than study estimate quality; or (4) they are not intended for estimating costs used in regulatory development.

Reason (3) applies to the *Cost Digest*, for example, as this report, designed for use by non-technical personnel, contains procedures for making "order-of-magnitude" estimates (± 30% accuracy or worse). *A Standard Procedure*, conversely, was primarily intended for estimating costs for R&D cases (e.g., demonstration projects), where some site-specific data are available. Further, although the latter report contains a thorough list of equipment installation factors, it contains few equipment *costs*. The report, *Evaluation of Control Technologies*, used data and estimating procedures from the 1978 *Manual* to provide sound generalized procedures for estimating thermal and catalytic incinerator costs. The third edition of the Manual updates and expands this information.

Finally, the second edition of the Manual (published December 1978), was one of the earliest of its kind. It has been extensively used in Agency regulatory development efforts. Accordingly, the Manual's role in the speciality of air pollution control system cost estimating is both unique and secure.

References for Section 1:

1. Neveril, R. B. (GARD, Inc.) *Capital and Operating Costs for Selected Air Pollution Control Systems*, EPA, Office of Air Quality Planning and Standards, Economic Analysis Branch, December 1978 (EPA 450/5-80-002).

2. DeWolf, Glenn, et al. (Radian, Inc.) *The Cost Digest: Cost Summaries of Selected Environmental Control Technologies*. EPA, ORD, Office of Environmental Engineering and Technology, October 1984 (EPA-600/884-010).

3. Vhl, Vincent W. *A Standard Procedure for Cost Analysis of Pollution Control Operations*, Volumes I and II. EPA, ORD, Industrial Environmental Research Laboratory, June 1979 (EPA-600/8-79-018a).

4. Katari, Vishnu (Pacific Environmental Services) *Evaluation of Control Technologies for Hazardous Air Pollutants*. EPA, Air and Energy Engineering Research Laboratory, ORD, October 1985.

Section 2

MANUAL ESTIMATING METHODOLOGY

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This section presents a methodology that will enable the user, having knowledge of the source being controlled, to produce study-level cost estimates for a control system to control that source. The methodology, which applies to each of the control systems included in this Manual, is general enough to be used with other "add-on" systems as well. Further, the methodology may also be applicable to estimating costs of fugitive emission controls and of other nonstack abatement methods.

Before presenting this methodology in detail, we should first discuss the various kinds of cost estimates and then define the cost categories and engineering economy concepts employed in making the estimates.

2.1 Types of Cost Estimates

As noted above, the costs and estimating methodology in this Manual are directed toward the "study" estimate, of ± 30% accuracy. According to Perry's Chemical Engineer's Handbook, a study estimate is ".... used to estimate the economic feasibility of a project before expending significant funds for piloting, marketing, land surveys, and acquisition ... [However] it can be prepared at relatively low cost with minimum data."(1) Specifically, to make a study estimate, the following must be known:

- Location of the source within the plant;
- Rough sketch of the process flow sheet (i.e., the relative locations of the equipment in the system);
- Preliminary sizes of, and material specifications for, the system equipment items;

- Approximate sizes and types of construction of any buildings required to house the control system;
- Rough estimates of utility requirements (e.g., electricity);
- Preliminary flow sheet and specifications for ducting and piping;
- Approximate sizes of motors required.(1)

In addition, an estimate of the labor hours required for engineering and drafting is needed, as the accuracy of an estimate (study or otherwise) is highly dependent on the amount of engineering work expended on the project.

There are, however, four other types of estimates, three of which are more accurate than the study estimate. These are:(1)

- "Order-of-magnitude"--"a rule of-thumb procedure applied only to repetitive types of plant installations for which there exists good cost history". Its accuracy is >±30%. (However, according to Perry's, "...no limits of accuracy can safely be applied to it.") The sole input required for making this level of estimate is the control system's capacity (often measured by the maximum volumetric flowrate of the gas passing through the system). So-called "six-tenths factor" estimates (not to be confused with factored estimates) are examples of this type.
- "Scope"/"Budget authorization"/"Preliminary". This estimate, nominally of ± 20% accuracy, requires more detailed knowledge than the study estimate regarding the site, flow sheet, equipment, buildings, etc. In addition, rough specifications for the insulation and instrumentation are also needed.

- "Project control"/"Definitive" estimates, accurate to within ± 10%, require yet more information than the scope estimates, especially concerning the site, equipment, and electrical requirements.
- "Firm"/"Contractor's"/"Detailed". This is the most accurate (± 5%) of the estimate types, requiring complete drawings, specifications, and site surveys. Further, "[t]ime seldom permits the preparation of such estimates prior to an approval to proceed with the project."(1)

For the purposes of regulatory development, study estimates have been found to be acceptable, as they represent a compromise between the less accurate "order-of-magnitude" and the more accurate estimate types. The former are too imprecise to be of much value, while the latter are not only very expensive to make, but require detailed site and process-specific knowledge that most Manual users will not have available to them.

2.2 Cost Categories Defined

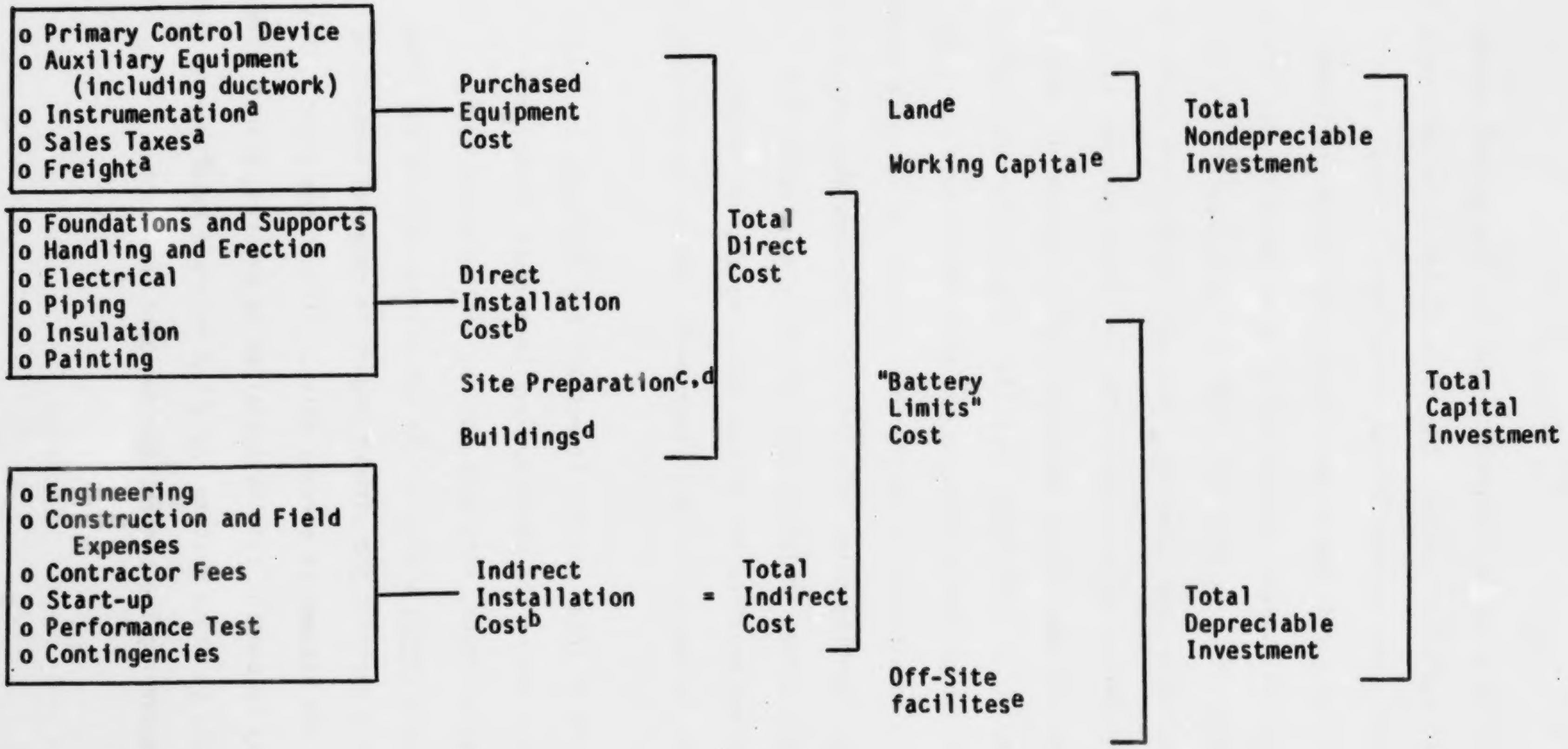
The names given certain "categories" of costs and what they contain vary considerably throughout the literature. Certain words like "capital cost" can have vastly different meanings, which can often lead to confusion, even among cost estimators. To avoid this confusion and, at the same time, provide uniformity in the Manual, basic terms are defined in this Section and will be used throughout. The terminology used is adapted from that of the American Association of Cost Engineers(2). Although it has been developed for general use, it is readily adaptable to air pollution control system costing.

2.2.1 Elements of Total Capital Investment

First, two general kinds of costs are estimated, total capital investment (TCI) and total annual cost (TAC). The total capital investment includes all costs required to purchase equipment needed for the control system (termed "purchased equipment" costs), the costs of labor and materials for installing that equipment (termed "direct installation" costs), site preparation, building costs, and certain other costs which are termed "indirect installation" costs. Direct installation costs include costs for foundations and supports, erecting and handling the equipment, electrical work, piping, insulation, and painting. Indirect installation costs include such costs as engineering costs; construction and field expenses (i.e., costs for construction supervisory personnel, office personnel, rental of temporary offices, etc.); contractor fees (for construction and engineering firms involved in the project); start-up and performance test costs (to get the control system running and to verify that it meets performance guarantees); and contingencies. Contingencies is a catch-all category that covers unforeseen costs that may arise, including (but certainly not limited to) "... possible redesign and modification of equipment, escalation increases in cost of equipment, increases in field labor costs, and delays encountered in start-up."(2)

These elements of total capital investment are displayed in Figure 2-1. Note that the sum of the purchased equipment cost, direct and indirect installation costs, site preparation, and buildings costs comprise the battery limits estimate. By definition, this is the total estimate "... for a specific job without regard to required supporting facilities which are assumed to already exist..."(2) at the plant. This would mainly apply to control systems installed in existing plants, though it could also apply to those systems installed in new plants when no special facilities for supporting the control system would be required.

Figure 2-1 Elements of Total Capital Investment


o Primary Control Device
o Auxiliary Equipment (including ductwork)
o Instrumentation[a]
o Sales Taxes[a]
o Freight[a]
Purchased Equipment Cost
o Foundations and Supports
o Handling and Erection
o Electrical
o Piping
o Insulation
o Painting
Direct Installation Cost[b]
Site Preparation[c,d]
Buildings[d]
Total Direct Cost
o Engineering
o Construction and Field Expenses
o Contractor Fees
o Start-up
o Performance Test
o Contingencies
Indirect Installation Cost[b]
=
Total Indirect Cost
Land[e]
Working Capital[e]
"Battery Limits" Cost
Off-Site facilites[e]
Total Nondepreciable Investment
Total Depreciable Investment
Total Capital Investment

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[a] These costs are factored from the sum of the control device and auxiliary equipment costs.
[b] These costs are factored from the purchased equipment cost.
[c] Usually required only at "grass roots" installations.
[d] Unlike the other direct and indirect costs, costs for these items are not factored from the purchased equipment cost. Rather, they are sized and costed separately.
[e] Normally not required with add-on control systems.

Where required, these supporting facilities would encompass units to produce steam, electricity, and treated water; laboratory buildings, railroad spurs, roads, and the like. It is unusual, however, for a control system to have one of these units (e.g., a power plant) dedicated to it. The system needs are rarely that great. However, it may be necessary--especially in the case of control systems installed in new or "grass roots" plants--for extra capacity to be built into the site generating plant to service the system. (A venturi scrubber, which often requires large amounts of electricity, is a good example of this.) It is customary for the utility costs to be charged to the project as operating costs at a rate which covers both the investment and operating costs for the utility.

As Figure 2-1 shows, there are two other costs which may be included in the total capital investment for a control system. These are "working capital", and "land". The first of these, working capital, is a fund set aside to cover the initial costs of fuel, chemicals, and other materials, as well as labor and maintenance. It usually does not apply to control systems, for the quantities of utilities, materials, labor, etc., they require are usually small. (An exception might be an oil-fired thermal incinerator, where a small supply (e.g., 30-day) of distillate fuel would have to be available during its initial period of operation.)

Land may also be required. But, since most add-on control systems take up very little space (a quarter-acre or less) this cost would be relatively small. (Certain control systems, such as those used for flue gas desulfurization, require larger quantities of land for the process equipment, chemicals storage, and waste disposal.)

Note also in Figure 2-1 that the working capital and land are "nondepreciable" expenses. In other words, these costs are "recovered" when the control system reaches the end of its useful life (generally in 10 to 20 years). Conversely, the other capital costs are "depreciable", in that they cannot be recovered and are included in the calculation of the income tax credit and depreciation allowance, whenever such taxes are considered in a cost analysis. (In the Manual methodology, however, income taxes are not considered. See Section 2.3.)

Notice that when 100% of the system costs are depreciated, no salvage value is taken for the system equipment at the conclusion of its useful life. This is a reasonable assumption for add-on control systems, as most of the equipment, which is designed for a specific source, cannot be used elsewhere without modifications. Even if it were reusable, the cost of disassembling the system into its components could be as high (or higher) than the salvage value.

2.2.2 Elements of Total Annual Cost

The Total Annual Cost (TAC) for control systems is comprised of three elements: "direct" costs, (DC) "indirect" costs, (IC) and "recovery credits" (RC), which are related by the following equation:

$$TAC = DC + IC - RC \qquad (2\text{-}1)$$

Clearly, the basis of these costs is one year, as this period allows for seasonal variations in production (and emissions generation) and is directly usable in profitability analyses. (See Section 2.3.)

Direct costs are those which tend to be proportional or partially proportional to the quantity of exhaust gas processed by the control system per unit time. These include costs for raw materials, utilities (steam, electricity, process and cooling water, etc.), waste treatment and disposal, maintenance materials, replacement parts, and operating, supervisory, and maintenance labor. Of these direct costs, costs for raw materials, utilities, and waste treatment and disposal are variable, in that they tend to be a direct function of the exhaust flowrate. That is, when the flowrate is at its maximum rate, these costs are highest. Conversely, when the flowrate is zero, so are the costs.

Semivariable direct costs are only partly dependent upon the exhaust flowrate. These include all kinds of labor, maintenance materials, and replacement parts. Although these costs are a function of the gas flowrate, they are not linear functions. Even while the control system is not operating, some of the semivariable costs continue to be incurred.

Indirect, or "fixed", annual costs are those whose values are totally independent of the exhaust flowrate and, in fact, would be incurred even if the control system were shut down. They include such categories as overhead, property taxes, insurance, and capital recovery.

Finally, the direct and indirect annual costs are offset by recovery credits, taken for materials or energy recovered by the control system, which may be sold, recycled to the process, or reused elsewhere at the site. These credits, in turn, must be offset by the costs necessary for their purification, storage, transportation, and any other costs required to make them reusable or resalable. Great care and judgement must be exercised in assigning values to recovery credits since materials recovered may be of small quantity or of doubtful purity, resulting in their having less value than virgin material.

The various annual costs and their interrelationships are displayed in Figure 2-2. A more thorough description of these costs and how they may be estimated is given in Section 2.4.

2.3 Engineering Economy Concepts

As mentioned previously, the estimating methodology presented in Section 2.4 rests upon the notion of the "factored" or study estimate. However, there are other concepts central to the cost analyses which must be understood. These are (1) the time value of money, (2) cash flow, and (3) annualization.

2.3.1 Time Value of Money

The "time value of money" is based on the truism that "...a dollar now is worth more than the prospect of a dollar... at some later date."(3) A measure of this value is the interest rate which "...may be thought of as the return obtainable by the productive investment of capital."(3)

2.3.2 Cash Flow

During the lifetime of a project, various kinds of cash expenditures are made and various incomes are received. The amounts and timing of these expenditures and incomes constitute the cash flows for the project. In control system costing it is normal to consider expenditures (negative cash flows) and unusual to consider income (positive cash flows), except for product or energy recovery income. By the simplifying convention recommended by Grant, Ireson, and Leavenworth(4), each annual expenditure (or payment) is considered to be incurred at the end of the year, even though the payment will probably be made sometime during the year in question. (The error introduced by this assumption is minimal, however.) Figure 2-3, which shows three hypothetical cash flow "diagrams", illustrates these end-of-year payments. In these diagrams, "P" represents the capital investment,

Figure 2-2 Elements of Total Annual Cost

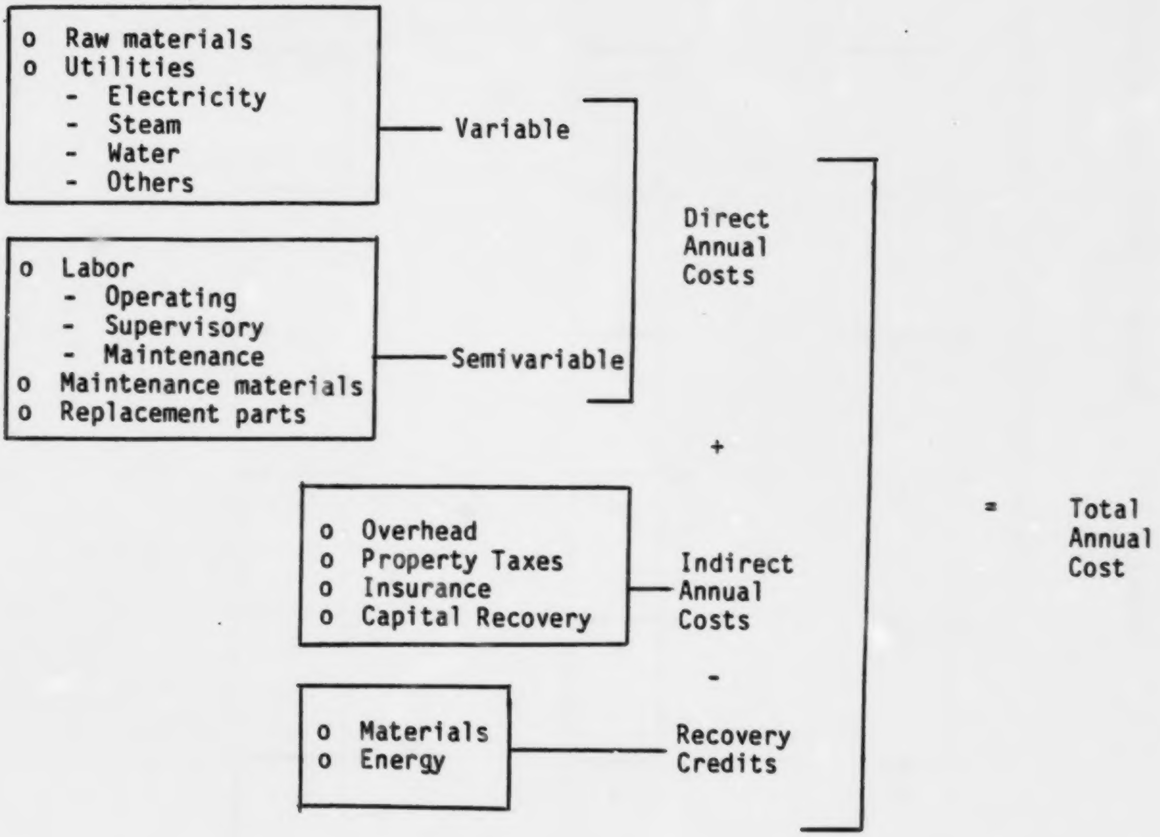
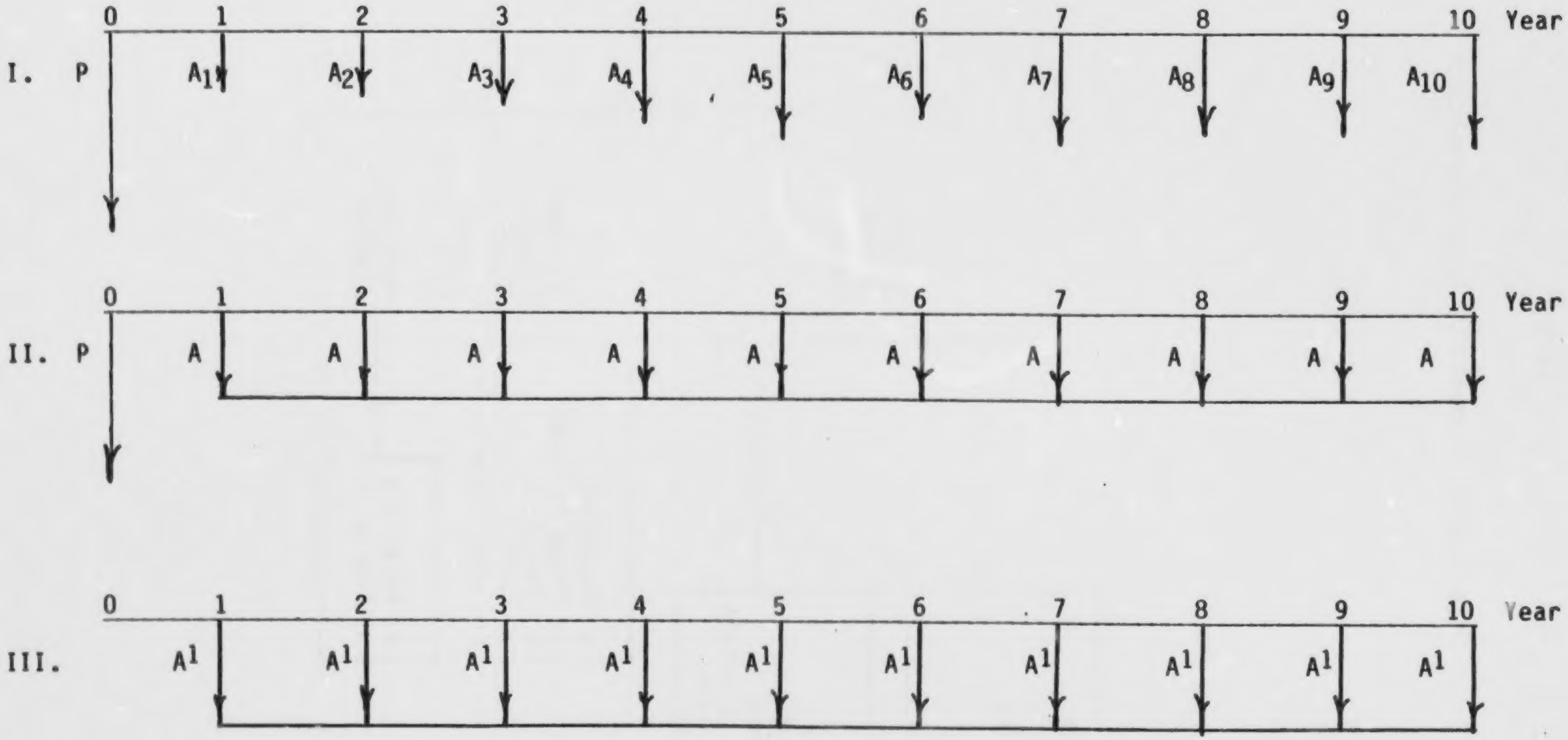

o Raw materials
o Utilities
- Electricity
- Steam
- Water
- Others
Variable
o Labor
- Operating
- Supervisory
- Maintenance
o Maintenance materials
o Replacement parts
Semivariable
Direct Annual Costs
+
o Overhead
o Property Taxes
o Insurance
o Capital Recovery
Indirect Annual Costs
-
o Materials
o Energy
Recovery Credits
=
Total Annual Cost

Figure 2-3 Hypothetical Cash Flow Diagrams

All Values Are Constant Year (Real) Dollars


I.
P
0
1
2
3
4
5
6
7
8
9
10
Year
A_1
A_2
A_3
A_4
A_5
A_6
A_7
A_8
A_9
A_{10}
II.
P
A
III.
A^1

while the A's denote the end-of-year annual payments. Note that in all diagrams, the cash flows are in "constant" ("real") dollars, meaning that they do not reflect the effects of inflation. Also note that in the top diagram (I), the annual payments are different for each year. (These represent the control system annual costs (exclusive of capital recovery) described in Section 2.2.) In reality, these payments would be different, as labor and maintenance requirements, labor and utility costs, etc., would vary from year to year. A generally upward trend in annual costs would be seen, however.

In diagram II, these fluctuating annual payments have been converted to equal payments. This can be done by calculating the sum of the present values of each of the annual payments shown in diagram I and annualizing the total net present value to equivalent equal annual payments via a capital recovery factor. (See discussion in the following paragraphs and in Section 2.3.3.) Alternatively, it is adequate to choose a value of A equal to the sum of the direct and indirect annual costs estimated for the first year. This assumption is in keeping with the overall accuracy of study estimates and allows for easier calculations.

Finally, notice diagram III. Here, the annual costs (A^1) are again equal, while the capital investment (P) is missing. Put simply, P has been incorporated into A^1 so that A^1 reflects not only the various annual costs but the investment as well. This was done by introducing another term, the capital recovery factor (CRF), defined as follows: "when multiplied by a present debt or investment , [the CRF] gives the uniform end-of-year payment necessary to repay the debt or investment in n years with interest rate i."(5) The product of the CRF and the investment (P) is the capital recovery cost (CRC):

$$CRC = CRF \times P \qquad (2\text{-}2)$$

where:

$$\mathrm{CRF} = \frac{i(1 + i)^n}{(1 + i)^n - 1} \qquad (2\text{-}3)$$

Therefore, A^1 is the sum of A and the CRC, or:

$$A^1 = A + \mathrm{CRF} \times P \qquad (2\text{-}4)$$

In this context, "n" is the control system economic life, which, as stated above, typically varies from 10 to 20 years. The interest rate ("i") used in this Manual is a real rate of 10% (annual). This value is used in most of the OAQPS cost analyses and is in keeping with current EAB guidelines(6) and the Office of Management and Budget recommendation for use in regulatory analyses.(7)

It may be helpful to illustrate the difference between "real" and "nominal" interest rates. The mathematical relationship between them is straightforward:(6)

$$(1 + i_n) = (1 + i)(1 + r) \qquad (2\text{-}5)$$

where:

i_n, i = the annual nominal and real interest rates, respectively

r = the annual inflation rate

Clearly, the "real" rate does not consider inflation and is in keeping with the expression of annual costs in constant (i.e., real) dollars.

EAB guidelines also recommend the exclusion of income tax considerations from cost analyses.(6) Not only does this simplify the analysis, but it also allows for the calculation of the "economic" costs of air pollution control--i.e., the true cost to society. Income taxes generally represent transfer payments from one segment of society to another and as such are not properly part of the economic costs.

2.3.3 Annualization Methods

The above method of "smoothing out" the investment into equal end-of-year payments, is termed the "equivalent uniform annual cash flow" (EUAC) method.(8) In addition to its inherent simplicity, this method is very useful when comparing the costs of two or more alternative control systems (i.e., those which are designed to control the same source to an equivalent degree). In fact, the EUAC's--or simply the total annual costs--of two competing systems may be compared even if both the systems have different economic lives, say 10 and 20 years. We recommend that the EUAC method be used for control cost work unless particular circumstances preclude its use.

Comparisons of systems with differing economic lives cannot be made, however, using the other two annualization (i.e., profitability analysis) methods--"present worth" and "internal rate of return". The "present worth" (or "discounted cash flow") method involves the "discounting" of all cash flows occuring after year "0" (i.e., the system startup date) back to year "0". These cash flows are discounted by multiplying each by a "discount factor", $\frac{1}{(1+i)^m}$, where "m" is the number of years from year 0 to the year in which the cash flow is incurred. The sum of these discounted cash flows is then added to the capital investment to yield the "present worth" of the project. The alternative having the highest present worth would be selected (in control system costing this is usually a negative number). But when comparing the present worths of alternative systems, the system lifetimes must be equal for the comparison to be valid.(9)

The third annualization method, "internal rate of return" (IRR), is similar to the present worth method, in that it involves the discounting of a series of unequal cash flows. However, where with the PW method the interest rate, "i", is set beforehand, in the IRR method the interest rate is solved for (usually via trial-and-error) after arbitrarily setting the PW to zero. When comparing alternative systems, the one with the highest IRR is selected.(10) But again, the alternative systems compared must have equal economic lives.

2.4 Estimating Procedure

The estimating procedure used in the Manual consists of five steps: (1) obtaining the "facility parameters" and "regulatory options" for a given facility; (2) "roughing out" the control system design; (3) sizing the control system components; (4) estimating the costs of these individual components; and (5) estimating the costs (capital and annualized) of the entire system.

2.4.1 Facility Parameters and Regulatory Options

Obtaining the facility parameters and regulatory options involves not only assembling the parameters of the air pollution source (i.e., the quantity, temperature, and composition of the emission stream(s)), but also compiling data for the facility's operation. (Table 2-1 lists examples of these.) Note that two kinds of facility parameters are identified -- "intensive" and "extensive". The former are simply those variables whose values are independent of quantity or dimensions -- i.e., the "extent" of the system. Conversely, "extensive" parameters encompass all size-dependent variables, such as the gas volumetric flowrate.

Table 2-1 Facility Parameters and Regulatory Options (11)

FACILITY PARAMETERS

- o Intensive
  - Facility status (new or existing, location)
  - Gas characteristics (temperature, pressure, moisture content)
  - Pollutant concentration(s) and/or particle size distribution
- o Extensive
  - Facility capacity
  - Facility life
  - Gas flow rate
  - Pollutant emission rate(s)

REGULATORY OPTIONS

- o No control
- o "Add-on" devices
  - Emission limits
  - Opacity "
- o Process modifications
  - Raw material changes
  - Fuel substitution
- o Others
  - Coal desulfurization

Like the facility parameters, the "regulatory options" are usually specified by others. These options are ways to achieve a predetermined emission limit. They range from "no control" to maximum control technically achievable. The option provided will depend, firstly, on whether the emission source is a stack (point source), a process leak ("process fugitives" source) or an unenclosed (or partly closed) area, such as a storage pile ("area fugitives" source). Stacks are normally controlled by "add-on" devices. As discussed above, this Manual will deal primarily with these add-on devices. (However, some of these devices can be used to control process fugitives in certain cases, such as a fabric filter used in conjunction with a building evacuation system.) Add-ons are normally used to meet a specified emission level, although in the case of particulate emissions, they may also be required to meet an opacity level.

2.4.2 Control System Design

Step 2 -- roughing out the control system design -- first involves deciding what *kinds* of systems will be priced (a decision that will depend on the pollutants to be controlled, gas stream conditions, and other factors), and what auxiliary equipment will be needed. When specifying the auxiliary equipment, several questions need to be answered:

- What type of hood (if any) will be needed to capture the emissions at the source?
- Will a fan be needed to convey the exhaust through the system?
- Is a cyclone or another pre-cleaner needed to condition the exhaust before it enters the control device?
- Will the captured pollutants be disposed of or recycled? How will this be done?

- Can the on-site utility capacity (e.g., electricity) accomodate the added requirements of the control system?

The kinds of auxiliary equipment selected will depend on the answers to these and other site-specific questions. However, regardless of the source being controlled, each system will likely contain, along with the control device itself, the following auxiliaries:

- Hood, or other means for capturing the exhaust;
- Ductwork, to convey the exhaust from the source, to, through, and from the control system;
- Fan system (fan, motor, starter, inlet/outlet dampers, etc.), to move the exhaust through the system;
- Stack, for dispersing the cleaned gas into the atmosphere.

2.4.3 Sizing the Control System

Once the system components have been selected, they must be sized. Sizing is probably the most critical step, because the assumptions made in this step will more heavily influence the capital investment than any other. Before discussing how to size equipment, we need to define the term. For the purposes of this *Manual*, "sizing" is the calculation (or estimation) of certain "critical" design parameters for a control device against which the purchased cost of that device is most accurately correlated. For instance, the purchased cost of an electrostatic precipitator (ESP) is most often correlated with its collecting area. This, in turn, is a function of the exhaust volumetric flowrate, the overall collection efficiency and the empirically-determined drift velocity, the ESP "critical" parameter. (Table 2-2 lists examples of these parameters.)

Table 2-2 Examples of Typical Control Device Parameters(11)

GENERAL

- Material of construction: e.g., carbon steel
- Insulated? Yes
- Economic life: 20 yr
- Redundancy[a]: none

DEVICE-SPECIFIC

- Air-to-cloth ratio ("critical parameter"): 7.5 to 1
- Pressure drop: 6.0 in w.g. (inches water gauge)
- Construction: suction (vs. pressurized)
- Duty: continuous (vs. intermittent)
- Other features: dilution air port (for exhaust temperature regulation)

[a] Refers to whether there are any extra equipment items installed (e.g., fans) to function in case the basic item becomes inoperative, so as to avoid shutting down the entire system.

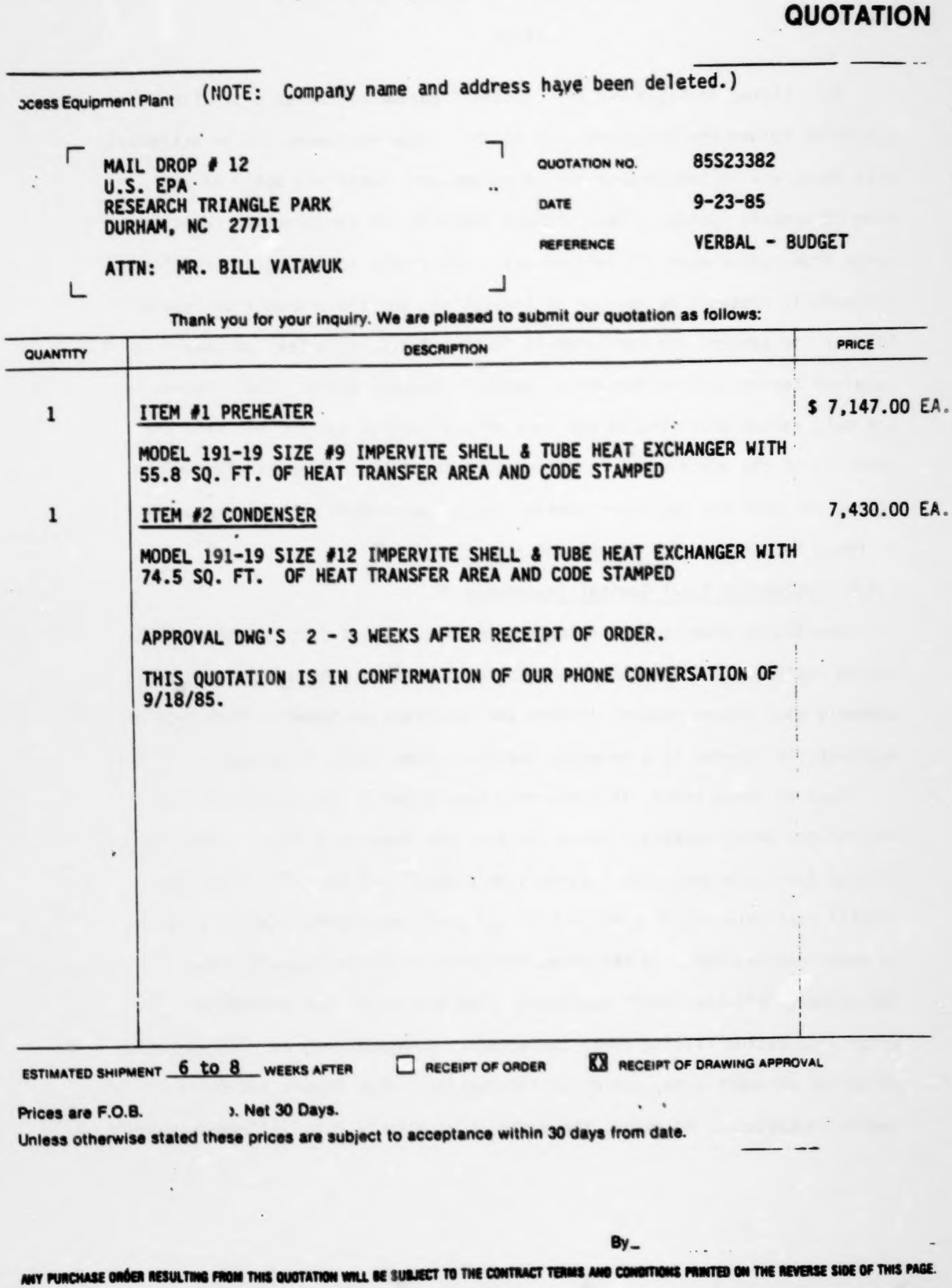
Also listed in Table 2-2 are "general" parameters which must also be specified before the purchased cost of the system equipment can be estimated. Note that, unlike the control device parameters, these may apply to any kind of control system. These include materials of construction (which may range from carbon steel to various stainless steels to fiberglass-reinforced polyester), presence or absence of insulation, and the economic or "useful" life of the system. As indicated in Section 2.3.2, this last parameter is required for estimating the annual capital recovery costs. The lifetime not only varies according to the type of the control system, but with the severity of the environment in which it is installed. (Representative values for this and the other control device parameters will be presented in those Sections of the *Manual* covering them.)

2.4.4 Estimating Total Capital Investment

The fourth step is estimating the total purchased cost of the control system equipment. These costs are available from this *Manual* for the most commonly used add-on control devices and auxiliary equipment. Each type of equipment is covered in a separate section. (See Table of Contents.)

Most of these costs, in turn, have been based on data obtained from control equipment vendors. There are over one hundred of these firms, many of whom fabricate and erect a variety of control systems.(12) They have readily available, current price lists of their equipment, usually indexed by model designation. If the items for which costs are requested are fabricated, "off-the-shelf" equipment, then the vendor can provide a written quotation listing their costs, model designations, date of quotation, estimated shipment date, and other information. (See Figure 2-4 for a sample quotation.) Moreover, the quote is usually "F.O.B." ("free-on-board")

Figure 2-4 Typical Vendor Quotation



QUOTATION

ocess Equipment Plant (NOTE: Company name and address have been deleted.)

MAIL DROP # 12
U.S. EPA
RESEARCH TRIANGLE PARK
DURHAM, NC 27711

ATTN: MR. BILL VATAVUK

QUOTATION NO. 85S23382
DATE 9-23-85
REFERENCE VERBAL - BUDGET

Thank you for your inquiry. We are pleased to submit our quotation as follows:

QUANTITY	DESCRIPTION	PRICE
1	ITEM #1 PREHEATER MODEL 191-19 SIZE #9 IMPERVITE SHELL & TUBE HEAT EXCHANGER WITH 55.8 SQ. FT. OF HEAT TRANSFER AREA AND CODE STAMPED	$ 7,147.00 EA.
1	ITEM #2 CONDENSER MODEL 191-19 SIZE #12 IMPERVITE SHELL & TUBE HEAT EXCHANGER WITH 74.5 SQ. FT. OF HEAT TRANSFER AREA AND CODE STAMPED	7,430.00 EA.
	APPROVAL DWG'S 2 - 3 WEEKS AFTER RECEIPT OF ORDER. THIS QUOTATION IS IN CONFIRMATION OF OUR PHONE CONVERSATION OF 9/18/85.	

ESTIMATED SHIPMENT 6 to 8 WEEKS AFTER ☐ RECEIPT OF ORDER ☒ RECEIPT OF DRAWING APPROVAL

Prices are F.O.B.). Net 30 Days.
Unless otherwise stated these prices are subject to acceptance within 30 days from date.

By

ANY PURCHASE ORDER RESULTING FROM THIS QUOTATION WILL BE SUBJECT TO THE CONTRACT TERMS AND CONDITIONS PRINTED ON THE REVERSE SIDE OF THIS PAGE.

the vendor, meaning that no taxes, freight, or other charges are included. However, if the items are not off-the-shelf, they must be custom fabricated or, in the case of very large systems, constructed on-site. In such cases, the vendor can still give quotations--but will likely take much longer to do so and may even charge for this service, to recoup the labor and overhead expenses of his estimating department.

As discussed in Section 2.2 in this Manual, the total capital investment is "factored" from the "purchased equipment cost", which in turn, is the sum of the base equipment cost (control device plus auxiliaries), freight, instrumentation, and sales tax. The values of these installation factors depend on the type of the control system installed and are, therefore, listed in the individual Manual sections dedicated to them.

The costs of freight, instrumentation, and sales tax are calculated differently from the direct and indirect installation costs. These items are "factored" also, but from the base equipment cost (F.O.B. the vendor(s)). But unlike the installation factors, these factors are essentially equal for all control systems. Values for these are as follows:

Cost	Range	Typical
Freight	0.01 - 0.10	0.05
Sales tax	0 - 0.08	0.03
Instrumentation	0.05 - 0.30	0.10

The range in freight costs reflects the distance between the vendor and the site. The lower end is typical of major U.S. metropolitan areas, while the latter would reflect freight charges to remote locations such as Alaska and Hawaii.(11) The sales tax factors simply reflect the range of local and state tax rates currently in effect in the U.S.(13)

The range of instrumentation factors is also quite large. For systems requiring only simple continuous or manual control, the lower factor would apply. However, if the control is intermittent and/or requires safety backup instrumentation, the higher end of the range would be applicable.(11) Finally, some "package" control systems (e.g., incinerators covered in Section 3) have built-in controls, whose cost is included in the base equipment cost. In those cases, the factor to use would, of course, be zero.

2.4.5 Retrofit Cost Considerations

The installation factors listed elsewhere in the Manual apply primarily to systems installed in new facilities. These factors must be adjusted whenever a control system is sized for, and installed in (i.e, "retrofitted") an existing facility. However, because the size and number of auxiliaries are usually the same in a retrofit situation, the total purchased cost of the control system would probably not be different from the new plant purchased cost. An exception is the ductwork cost, for in many retrofit situations exceptionally long duct runs are required to tie the control system into the existing process.

Each retrofit installation is unique; therefore, no general factors can be developed. Nonetheless, some general information can be given concerning the kinds of system modifications one might expect in a retrofit:

1. Auxiliaries. Again, the most important component to consider is the ductwork cost. In addition, to requiring very long duct runs, some retrofits require extra tees, elbows, dampers, and other fittings.
2. Handling and Erection. Because of a "tight fit", special care may need to be taken when unloading, transporting, and placing the

equipment. This cost could increase significantly if special means (i.e., helicopters) are needed to get the equipment on roofs or to other inaccessible places.

3. Piping, Insulation, and Painting. Like ductwork, large amounts of piping may be needed to tie in the control device to sources of process and cooling water, steam, etc. Of course, the more piping and ductwork required, the more insulation and painting will be needed.
4. Site Preparation. Unlike the other categories, this cost may actually decrease, for most of this work would have been done when the original facility was built.
5. Facilities. Conceivably, retrofit costs for this category could be the largest. For example, if the control system requires large amounts of electricity (e.g., a venturi scrubber), the facility's power plant may not be able to service it. In such cases, the facility would have to purchase the additional power from a public utility, expand its power plant, or build another one. In any case, the cost of electricity supplied to that control system would likely be higher than if the system were installed in a new facility where adequate provision for its electrical needs would have been made.
6. Engineering. Designing a control system to fit into an existing plant normally requires extra engineering, especially when the system is exceptionally large, heavy, or utility-consumptive. For the same reasons, extra supervision may be needed when the installation work is being done.

7. Lost Production. This cost is incurred whenever a retrofit control system cannot be tied into the process during normally scheduled maintenance periods. Then, part or all of the process may have to be temporarily shut down. The revenue lost during this shutdown period is a bonafide retrofit expense.
8. Contingency. Due to the uncertain nature of retrofit estimates, the contingency (i.e., uncertainty) factor in the estimate should be increased.

From the above points, it is apparent that some or most of these installation costs would increase in a retrofit situation. However, there may be other cases where the retrofitted installation cost would be less than the cost of installing the system in a new plant. This could occur when one control device, say an ESP, is being replaced by a more efficient unit--a baghouse, for example. The ductwork, stack, and other auxiliaries for the ESP may be adequate for the new system, as perhaps would the support facilities (power plant, etc.).

2.4.6 Estimating Annual Costs

Determining the total annual cost is the last step in the estimating procedure. As mentioned in Section 2.2 the TAC is comprised of three components -- direct and indirect annual costs and recovery credits. Unlike the installation costs, which are "factored" from the purchased equipment cost, annual cost items are usually computed from known data on the system size and operating mode, as well as from the facility and control device parameters.

Following is a more detailed discussion of the items comprising the total annual cost. (Values/factors for these costs are also given in the sections for the individual devices.)

2.4.6.1 Raw Materials

Raw materials are generally not required with control systems. Exceptions would be chemicals used in absorbers or venturi scrubbers as absorbents or to neutralize acidic exhaust gases (e.g., hydrochloric acid). Chemicals may also be required to treat wastewater discharged by scrubbers or absorbers before releasing it to surface waters. But, these costs are only considered when a wastewater treatment system is exclusively dedicated to the control system. In most cases, a pro-rata waste treatment charge is applied. (See section 2.4.6.5.)

Quantities of chemicals required are calculated via material balancies, with an extra 10 to 20% added for miscellaneous losses. Costs for chemicals are available from the Chemical Marketing Reporter and similar publications.

2.4.6.2 Operating Labor

The amount of labor required for a system depends on its size, complexity, level of automation, and operating mode (i.e., batch or continuous). The labor is usually figured on an hours-per-shift basis. As a rule, though, data showing explicit correlations between the labor requirement and capacity are hard to obtain. A typical correlation is logarithmic:

$$\frac{L_2}{L_1} = \left(\frac{V_2}{V_1}\right)^y \qquad (2\text{-}6)$$

where: L_1, L_2 = labor requirements for systems 1 and 2

V_1, V_2 = capacities of systems 1 and 2 (as measured by the gas flow rate, for instance)

y = 0.2 to 0.25 (typically)(14)

The exponent in equation (2-6) can vary considerably, however. Conversely, in many cases, the amount of operator labor required for a system will be approximately the same regardless of its size.

A certain amount must be added to operating labor to cover supervisory requirements. Fifteen per cent of the operating labor requirement is representative.(15)

To obtain the annual labor cost, multiply the operating and supervisory labor requirements by the respective wage rates (in $/hr) and the system operating factor (number of hours per year the system is in operation). The wage rates also vary widely, depending upon the source category, geographical location, etc. These data are tabulated and periodically updated by the U.S. Department of Labor, Bureau of Labor Statistics, in its _Monthly Labor Review_ and in other publications. Finally, note that these are _base_ labor rates, which do _not_ include payroll and plant overhead. (See overhead discussion below.)

2.4.6.3 Maintenance

Maintenance labor is calculated in the same way as operating labor and is influenced by the same variables. The maintenance labor rate, however, is normally higher than the operating labor rate, mainly because more skilled personnel are required. A 10% wage rate premium is typical.(15)

Further, there are expenses for maintenance materials -- oil, other lubricants, duct tape, etc., and a host of small tools. Costs for these items can be figured individually, but since they are normally so small, they are typically factored from the maintenance labor. Reference 15 suggests a factor of 100% of the maintenance labor.

2.4.6.4 Utilities

This cost category covers many different items, ranging from electricity to compressed air. Of these, only electricity is common to all control devices, where fuel oil and natural gas are generally used only by incinerators; water and water treatment, by venturi scrubbers, quenchers, and spray chambers; steam, by carbon adsorbers; and compressed air, by pulse-jet fabric filters.

Techniques and factors for estimating utility costs for specific devices are presented in their respective sections. However, because nearly every system requires a fan to convey the exhaust gases to and through it, a general expression for computing the fan electricity cost (C_e) is given here:(11)

$$C_e = \frac{0.746\ Q\ \Delta P\ s\ \theta\ P_e}{6356\ \eta} \qquad (2\text{-}7)$$

where:

Q = gas flowrate (actual ft^3/min)

ΔP = pressure drop through system (inches of water, gauge) (Values for ΔP are given in the sections covering the equipment items.)

s = specific gravity of gas relative to air (1.000, for all practical purposes)

θ = operating factor (hr/yr)

η = combined fan and motor efficiency (usually 0.60 to 0.70)

P_e = electricity cost ($/kwhr).

A similar expression can be developed for calculating pump motor electricity requirements.

2.4.6.5 Water Treatment and Disposal

Though often overlooked, there can be a significant cost associated with treating and/or disposing of waste material captured by a control system that neither can be sold nor recycled to the process.

Liquid waste streams, such as the effluent from a venturi scrubber, are usually processed before being released to surface waters. The type and extent of this processing will, of course, depend on the characteristics of the effluent. For example, the waste can first be sent to one (or more) clarifiers, for coagulation and removal of suspended solids. The precipitate from the clarifier is then conveyed to a rotary filter, where most of the liquid is removed. The resulting filter cake is then disposed of, via land filling, for example.

The annual cost of this treatment is relatively high--\$1.00 to \$2.00/thousand gallons treated or more.(16) The solid waste disposal costs (via land filling, for example) typically would add another \$20 to \$30/ton disposed of.(17) This, however, would not include transportation to the disposal site. More information on these technologies and their costs is found in References (16) and (17).

2.4.6.6 Replacement Parts

This cost is computed separately from maintenance, because it is a large expenditure, incurred one or more times during the useful life of a control system. This category includes such items as carbon (for carbon adsorbers), bags (for fabric filters) and catalyst (for catalytic incinerators), along with the labor for their installation.

The annual cost of the replacement materials is a function of the initial parts cost, the parts replacement labor cost, the life of the parts, and the interest rate, as follows:

$$CRC_p = (C_p + C_{pl})CRF_p \qquad (2\text{-}8)$$

where: CRC_p = capital recovery cost of replacement parts (\$/yr)

C_p = initial cost of replacement parts, including taxes and freight (\$)

C_{pl} = cost of parts replacement labor (\$)

CRF_p = capital recovery factor (defined in Section 2.3).

In the Manual methodology, replacement parts are treated the same as any other investment, in that they are also considered an expenditure that must be amortized over a certain period. Also, the useful life of the parts (typically 2 to 5 years) is generally less than the useful life of the rest of the control system.

Replacement part labor will vary, depending upon the amount of the material, its workability, accessibility of the control device, and other factors.

2.4.6.7 Overhead

This cost is easy to calculate, but often difficult to comprehend. Much of the confusion surrounding overhead is due to the many different ways it is computed and to the several costs it includes, some of which may appear to be duplicative.

There are, generally, two categories of overhead, payroll and plant. Payroll overhead includes expenses directly associated with operating, supervisory, and maintenance labor, such as: workmen's compensation, Social Security and pension fund contributions, vacations, group insurance, and other fringe benefits. Some of these are fixed costs (i.e., they must be paid regardless of how many hours per year an employee works). Payroll overhead is traditionally computed as a percentage of the total annual labor cost (operating, supervisory, and maintenance).

Conversely, plant (or "factory") overhead account for expenses not necessarily tied to the operation and maintenance of the control system, including: plant protection, control laboratories, employee amenities, plant lighting, parking areas, and landscaping. Some estimators compute plant overhead by taking a percentage of all labor plus maintenance materials(18), while others factor it from the total labor costs alone.(19)

For "study" estimates, it is sufficiently accurate to combine payroll and plant overhead into a single indirect cost. This is done in this Manual. Also, overhead will be factored from the sum of all labor (operating, supervisory, and maintenance) plus maintenance materials, the approach recommended in reference 18. The factors recommended therein range from 50 to 70% (18) An average value of 60% is used in this Manual.

2.4.6.8 Property Taxes, Insurance, and Administrative Charges

These three indirect operating costs are factored from the system total capital investment, and typically comprise 1,1, and 2% of it, respectively. Taxes and insurance are self-explanatory. "Administrative charges" covers sales, research and development, accounting, and other home office expenses. (It should not be confused with plant overhead, however.) For simplicity, the three items are usually combined into a single, 4% factor. This value, incidentally, is standard in all OAQPS cost analyses.

2.4.6.9 Capital Recovery

As discussed in Section 2.3, the annualization method used in the Manual is the equivalent uniform annualized cost method. Recall that the cornerstone of this method is the capital recovery factor which, when multiplied by the total capital investment, yields the capital recovery cost. (See equation 2-2.)

However, whenever there are parts in the control system that must be replaced before the end of its useful life, equation 2-2 must be adjusted, to avoid double-counting.

That is:

$$CRC_s = CRF_s \, [TCI - (C_p + C_{pl})] \qquad (2\text{-}9)$$

where:

CRC_s = capital recovery cost for control system (\$/yr)

TCI = total capital investment for entire system (\$).

CRF_s = capital recovery factor for control system

The term (C_p+C_{pl}) accounts for the cost of those parts that would be replaced during the useful life of the control system and the labor for replacing them. Clearly, CRF_s and CRF_p will not be equal unless the control system and replacement part lives are equal.

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Section 3

THERMAL AND CATALYTIC INCINERATORS

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3.1 Process Description

3.1.1 Elements of Combustion

Fume incineration, a controlled oxidation process, is a technique used for destruction of vaporous volatile organic compound (VOC) emissions from industrial waste gases. In the process, the VOC content of waste gases reacts at high temperatures with oxygen to form carbon dioxide and water, while liberating heat. Three parameters: temperature, residence time (also referred to as "retention time" or "dwell time") and turbulence (the "three Ts") have an interrelated effect upon the final combustion performance. To achieve good oxidation rates and obtain release of the full heat content of the combustion products, emission effluents must be held for sufficient residence times at combustion temperatures 100°F or more above reported ignition temperatures. Further, turbulent flow conditions must be maintained in the incinerator for good mixing of the fuel combustion products with incoming effluents, so that a thorough homogeneity of combustion elements (VOC and oxygen) may be achieved. The ignition temperature of a given VOC is the minimum temperature at which the VOC reacts instantaneously with oxygen. Below this temperature (if the supply of heat were interrupted) combustion would slow down and gradually stop.

Typically, the rudimentary combustion parameters for an incinerator design are established empirically and no calculations are involved in their estimation. Of the 3 Ts, time and turbulence are fixed by incinerator design and air flow rate, and only temperature can be significantly controlled. The type of VOCs present in the waste gas usually dictates the combustion temperature for a required VOC destruction rate. Lower or higher combustion temperatures can be used to a limited extent, with corresponding variations in the residence time (e.g., an incinerator can be designed for a lower residence time but must operate at a correspondingly higher combustion temperature). The possibilities of varying these parameters are limited, because the combustion temperature must always be above the ignition temperature. Moreover, the cost savings realized from reducing the residence time (and, in turn, the incinerator size) generally are not large enough to offset the additional expenses (e.g., fuel) resulting from compensating increases in the combustion temperature.

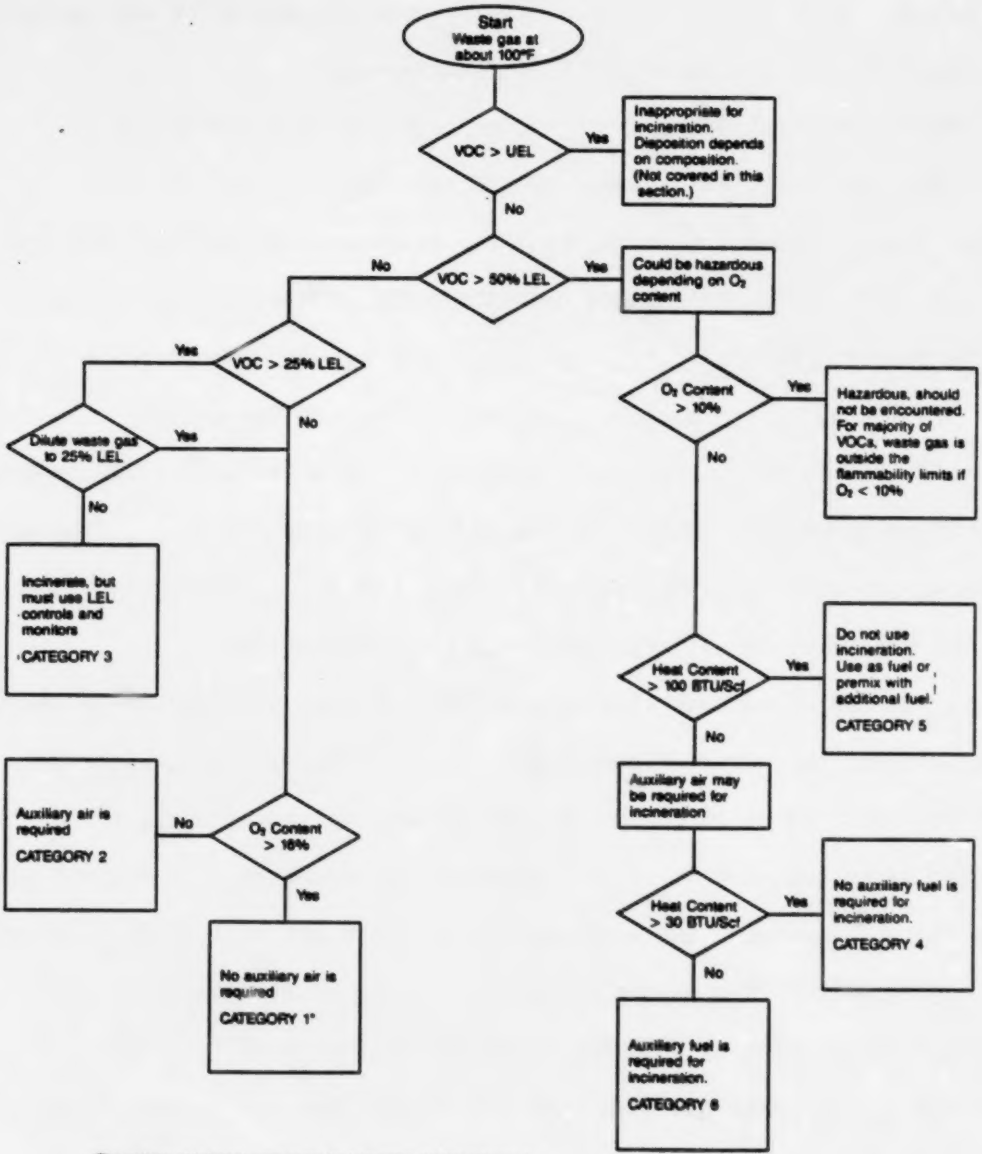
In general, the design of an incinerator system is based on providing the required amounts of: (1) heat to bring the waste gas to the oxidation temperature and (2) oxygen for complete combustion. The waste gas heat and oxygen contents determine, respectively, the auxiliary heat and oxygen requirements (i.e., the lower the waste gas heat and oxygen contents, the higher are the auxiliary requirements). Either natural gas or fuel oil can be used as an auxiliary heat source while ambient air is the usual source of oxygen. Natural gas is preferred and used in most incinerators. Oil-fired burners have limited turndown capabilities and require high maintenance. Depending upon the fuel availability and geographic location, some incinerators are designed for a dual fuel-burning capability (natural gas with standby oil). In practice, a flue gas oxygen content of at least 3 percent (by volume) is used empirically to insure that a sufficient amount of oxygen is available for combustion.

The waste gas heat content measures the potential heat of VOC that would be released as heat of combustion upon the oxidation of the VOC. The heat and VOC contents of a waste gas are proportional to, and can provide a measure of, each other. They are also inherently related to, and can be expressed empirically, in terms of the VOC property "flammability." Flammability is characterized by two limits: the lower explosive limit (LEL) and the upper explosive limit (UEL). These limits represent, respectively, the smallest and largest amounts of VOCs which, when mixed with air, will burn without a continuous application of external heat. Major reference books present VOC flammability (LEL and UEL) data. Table 3A-1 in Appendix 3A presents the flammability data for several VOCs commonly encountered in industrial waste gases. To avoid potential explosions, the VOC content of industrial waste gases released to the atmosphere is normally outside the flammability limits. The majority of waste gases contain low concentrations of VOCs. For safety reasons, their composition is typically limited to below 25 percent and will seldom exceed 50 percent of the LEL level. It has been empirically determined that, upon combustion, most VOCs release approximately 50 Btu/scf of waste gas, if their concentration is at 100 percent of the LEL. Therefore, the VOC content of a waste gas at less than 25 percent LEL, the safe upper limit for incineration, will have a potential heat content of less than 13 BTU/scf.

The LEL and heat content values of a given waste gas can also be related empirically to the temperature rise of the combustion flue gases. Upon combustion, each one percent of LEL of the VOC (i.e., a potential heat content of 0.50 BTU/scf of waste gas) can raise the temperature of one ft^3 of the waste gas by 27°F based on a heat capacity of 0.018 BTU/ft^3°F for dry air. These empirical relations can sometimes be used in making qualitative observations to determine if a given waste gas can be incinerated. They can also be used to make quick and approximate estimates of the auxiliary fuel requirement.

Most industrial waste gases incinerated are dilute mixtures of VOCs, air, and inert gas. Their VOC content is very low, and their oxygen content exceeds that required for combustion of both the VOC and the auxiliary fuel. If a waste gas with a VOC content over 25 percent LEL is encountered, it is diluted to below 25 percent LEL prior to incineration by adding outside air (thus increasing the waste gas volume flow rate to be treated). However, a waste gas with a VOC content from 25 to 50 percent LEL can be incinerated without adding dilution air, provided the waste gas VOC levels in the system are continuously monitored via LEL monitors, to satisfy fire protection regulations.

A low-oxygen content waste gas mixture of VOC, air, and inert gases could disrupt the burner flame stability. Therefore, when a waste gas with less than 13 to 16 percent oxygen content is incinerated, the portion of such waste gas used for fuel combustion is augmented with ambient air. In a few applications, the waste gas is an inert gas with a low VOC content and negligible or zero oxygen content. In such cases ambient air is provided for burning of both the waste gas VOC and auxiliary fuel. In rare cases, the waste gas is a rich VOC stream that can support combustion without auxiliary fuel. Such a rich VOC waste gas is treated as a fuel and burned, either premixed or oxygen free. This process is sometimes referred to as "direct flame incineration." Figure 3-1 provides a flow chart for categorizing a waste gas to determine its suitability for incineration. (For more information on waste gas characterization, see Appendix 3B.)


Start
Waste gas at about 100°F
VOC > UEL
Yes
Inappropriate for incineration. Disposition depends on composition. (Not covered in this section.)
No
VOC > 50% LEL
Yes
Could be hazardous depending on O_2 content
No
VOC > 25% LEL
Yes
Dilute waste gas to 25% LEL
Yes
No
No
Incinerate, but must use LEL controls and monitors
CATEGORY 3
O_2 Content > 10%
Yes
Hazardous, should not be encountered. For majority of VOCs, waste gas is outside the flammability limits if O_2 < 10%
No
Heat Content > 100 BTU/Scf
Yes
Do not use incineration. Use as fuel or premix with additional fuel.
CATEGORY 5
No
Auxiliary air may be required for incineration
Heat Content > 30 BTU/Scf
Yes
No auxiliary fuel is required for incineration.
CATEGORY 4
No
Auxiliary fuel is required for incineration.
CATEGORY 6
Auxiliary air is required
CATEGORY 2
No
O_2 Content > 16%
Yes
No auxiliary air is required
CATEGORY 1*

*The majority of industrial waste gases for incineration fall into category 1. Calculations for category 1 are addressed in the text and calculations for all categories (i.e. categories 1 through 6) are addressed in Appendix 3B.

Figure 3-1. Flow chart for categorization of a waste gas to determine its suitability for incineration and need for auxiliaries.

3.1.2 Types of Incinerators

Two types of incinerators are in use: thermal and catalytic. The combustion process, as well as equipment, design concept, and calculations, are essentially the same with both incinerator types. The only exception is that catalytic incineration is essentially a flameless combustion process, wherein a catalyst bed is used to initiate the combustion reaction at much lower temperatures.

Historically, the greatest reason for using a catalytic incinerator has been to reduce fuel consumption. However, because the application of either a thermal or a catalytic incinerator system without a heat exchanger (for recovery of flue gas heat) is now rare, the savings associated with catalytic incinerators are less significant. Also, this fuel economy for catalytic incinerators is partially offset by increased operating costs for maintenance (i.e., periodic cleaning and replacement of catalyst). Catalysts undergo a gradual loss of activity through thermal aging, fouling, and erosion of their surfaces. Certain poisonous contaminants, such as phosphorus, arsenic, antimony, lead, and zinc, also cause catalyst deactivation. Catalytic incineration is not recommended for waste gases containing significant concentrations of particulate matter (either organic or inorganic) that cannot be vaporized.

The catalysts used in catalytic incinerator systems for gaseous VOC control are usually precious or base metals or their salts, either supported on inert carriers, such as alumina or porcelain, or unsupported. Precious metal oxide catalysts are less brittle and more expensive than base metal types, and are used in lesser amounts per unit of waste gas volume. Of the precious metal oxide catalysts, platinum/palladium oxides are preferred. Others include rhodium, nickel, and gold.

Manganese dioxide is the most commonly used base metal oxide catalyst.

The types of VOCs present in the waste gas determine the operating temperature required in a catalytic unit. The more stable VOCs are generally the least reactive and require higher inlet catalytic operating temperatures. Methane is an example of a stable, low molecular weight compound that requires a relatively high catalytic conversion temperature, about 1,000°F. Hydrogen, on the other hand, is extremely reactive, having a conversion temperature of about 200°F. To achieve a VOC destruction efficiency of at least 90 percent, the catalytic ignition temperature for most hydrocarbons must be between 400 and 500°F. Higher temperatures are required to obtain higher VOC destruction efficiencies. When methane is present along with other less stable VOCs in a waste gas, preheat temperatures lower than 1,000°F can be used, because the less stable VOCs burn first and generate heat.

The auxiliary fuel requirement is the most significant operating expense of an incinerator. Further, the temperature rise of the combustion products due to the VOC heat release can be substantial. Therefore, to minimize auxiliary fuel expenses, a part of the heat from the incinerator flue gases is recovered. This heat is usually recovered in recuperative heat exchangers, in which the heat from the flue gases is exchanged with the waste gas. Both countercurrent and cross-flow types of heat exchangers are used for this purpose. When feasible, heat may be recovered indirectly, by producing low-pressure steam in a waste heat boiler. Further, heat recovery from the flue gases after primary heat recovery (PHR) can be achieved by employing secondary heat recovery (SHR) units. Of course, SHR can be economical, only if the secondary heat can be consumed on-site and cheaper heat sources are unavailable. The PHR and SHR combinations are most suitable for large installations where recovered secondary heat may be used on-site for water heating, air heating, and other purposes.

In general, a typical incinerator system may include the following components: (1) a fan to move the waste gas; (2) a fan to supply ambient air, if required; (3) a combustion unit (i.e., a refractory chamber with burner for thermal units, and a preheat chamber with burner and catalyst bed for catalytic units); (4) heat recovery equipment (optional, but almost always used); (5) controls, instrumentation, and control panel; (6) a stack; and (7) in the case of catalytic units, a filter/mixer to assure flow distribution, protect the catalyst bed from flame impingement, and remove noncombustible particulate matter. In addition, auxiliary equipment, such as ductwork, may be required in the system.

3.2 Design Procedure

3.2.1 Design and Operating Features

The minimum waste gas characteristics data needed to perform incinerator design calculations are the waste gas volume flow rate and temperature, and the VOC composition. The waste gas flow rate primarily determines the quantity of combustion flue gases generated which, in turn, dictates the size of an incinerator system. As discussed in Section 3.1, the waste gas composition determines the combustion air requirements. The waste gas and combustion temperatures, along with the waste gas volume flow rate and VOC content, determine the auxiliary heat requirement and the heat exchanger size. Finally, the types of VOC present in the waste gas determine the combustion temperature required in the incinerator for optimum oxidation.

In the incineration process, a waste gas is introduced to the combustion chamber, (or "preheat chamber," in the case of a catalytic incinerator), where the waste gas temperature is raised to the appropriate combustion temperature by burning auxiliary fuel. Because of the high combustion temperatures maintained, refractory chambers are used in thermal incinerators, while stainless steel or carbon steel chambers are used in catalytic incinerators. In thermal incinerators, the waste gas is heated and retained for 0.3 to 1.0 seconds in the combustion chamber at 100°F or more above the ignition temperature, which ranges from 1,000 to 1,400°F for most VOCs. At these temperatures, 95 to 99 percent of the VOCs in the waste gas are combusted. As discussed in Section 3.1, the resulting flue gases are exhausted via a stack to the atmosphere, after a part of their sensible heat is recovered via direct exchange with the incoming waste gas.

In catalytic incinerators, the waste gas temperature is typically raised in the preheat chamber to 500 to 600°F. This is above the catalytic ignition temperature of 400 to 550°F theoretically required for 90 percent destruction of most VOCs. (Thermal and catalytic ignition temperature data for VOCs are well documented in the literature. Table 3A-2 in Appendix A presents catalytic ignition temperatures for several common VOCs.) The thoroughly mixed gaseous effluents from the preheat chamber (where partial oxidation may occur) are subsequently passed through specially designed units containing catalyst elements, on the surface of which oxidation occurs at an accelerated rate at temperatures of 700-900°F--much lower than typical thermal incineration combustion chamber temperatures. As the linear velocity through the catalyst bed is high (600 to 1,200 ft/min.), the residence time is negligible.

For this reason residence time is rarely an important factor in the design of catalytic incinerators. The heat of reaction from the oxidation of the VOCs in the catalyst bed causes the gas temperature to increase as it passes across the catalyst bed. The amount of VOCs present in the waste gas determines the temperature increase in the catalyst bed.

The desired catalyst bed outlet temperature is typically 700 to 900°F. The maximum temperature to which the catalyst bed can be exposed continuously is limited to about 1,200°F. Therefore, the heat released from the combustion reaction and, accordingly, the VOC content of the waste gas are limited to about 20 percent LEL. (See Section 3.2.2. for more details.)

Depending upon the catalyst bed temperature swings (i.e., the frequency at which it is subjected to extreme temperature excursions), the operating and maintenance practices, and the particulate matter and specific catalyst poisons encountered, the catalyst would have an effective life of 2 to 10 years. The amount of catalyst required (measured by the standard hourly flue gas volume flow rate per unit volume of catalyst or the "space velocity," hr^{-1}), depends on the type of catalyst used, but increases with the required VOC destruction efficiency. The space velocity, incinerator gas velocity, and pressure drop used in the system design are determined experimentally.

Finally, as in thermal incinerators, the flue gases exiting the catalyst bed are exhausted to a stack, usually after heat exchange with the incoming waste gas.

3.2.2 Design Calculations

This section presents calculations for designing an incineration system to the level of detail required by a study cost estimate. Further, these calculations only apply to dilute mixtures of VOCs, air, and inert gas.

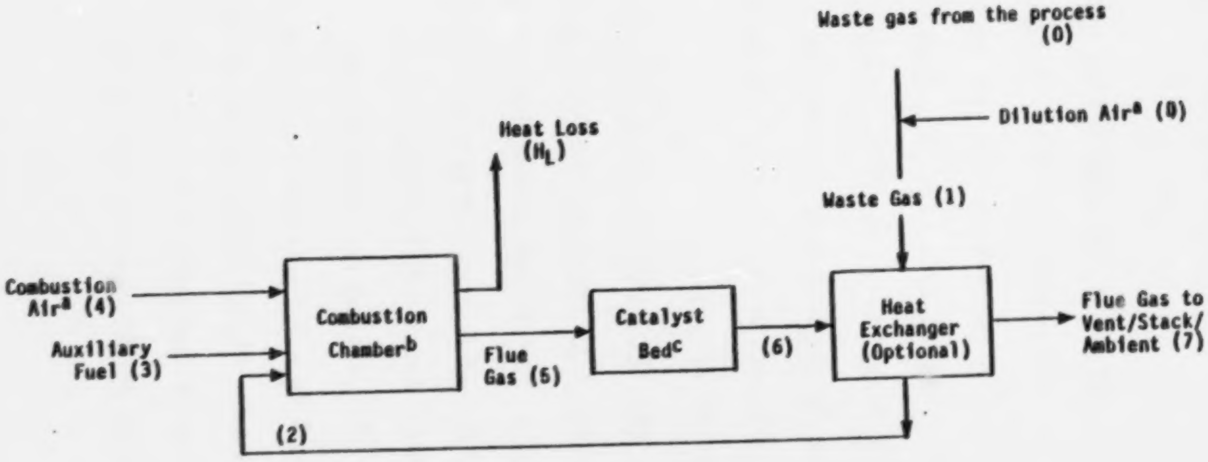
These mixtures are typical of most industrial waste gases (i.e., those cases which require no outside air for dilution or for combustion of fuel and waste gas VOC--Category 1 in Figure 3-1 and Table 3B-1, Appendix 3B). Appendix 3B presents a general procedure applicable to all types of waste gas compositions, including those requiring the use of outside air for combustion or dilution purposes (i.e., Categories 1 through 6 in Figure 3-1 and Table 3B-1, Appendix 3B).

Figure 3-2 is a simplified schematic of an incinerator system. A thermocouple in a thermal incinerator combustion chamber measures temperature, and appropriate control circuitry alters the rate of auxiliary fuel entering the incinerator to maintain the desired combustion temperature. In catalytic incinerators, thermocouples installed in the preheat chamber and catalytic bed perform the same function.

The incinerator design calculations for the waste gas cases where no outside air is added for either combustion or dilution and where the VOC content is low (i.e., Category 1) can be summarized by the following step-wise approach. The items that must be calculated or estimated in the design of any emission control system are those which determine the system size, performance, and capital and operating costs. In the case of incinerators, these items are: (1) the auxiliary fuel requirement and flue gas flow rate; (2) for catalytic incinerators, the amount of catalyst required; and (3) the pressure drop across the system.

Item 1. Calculate Auxiliary Fuel Requirement and Flue Gas Flow Rate.

The calculations of the auxiliary fuel requirement and the flue gas flow rate constitute an important incinerator design item. The flue gas flow rate determines the incinerator system size and, consequently, its capital cost. The auxiliary fuel can be a major operating cost for an incinerator system.


Waste gas from the process (0)
Dilution Air^a (0)
Waste Gas (1)
Heat Loss (H_L)
Combustion Air^a (4)
Auxiliary Fuel (3)
Combustion Chamber^b
Flue Gas (5)
Catalyst Bed^c
(6)
Heat Exchanger (Optional)
Flue Gas to Vent/Stack/ Ambient (7)
(2)

(a) When required.

(b) Referred to as Preheat Chamber in the case of catalytic incinerators.

(c) Included only in catalytic incinerators.

Figure 3-2. Schematic diagram of an incinerator system.

Calculations of the auxiliary fuel requirement and flue gas flow rate are considered as one item, because they are interrelated. The specified flue gas conditions determine the amount of auxiliary fuel consumed which, in turn, becomes a part of the flue gas. The necessary waste gas and combustion information to be compiled and the calculations to be performed are presented in the following steps:

Step 1. Identify the waste gas composition data. These include: the volume flow rate (Q_1), scfm; temperature (T_1), °F; VOC content, % LEL; and heat content (h_1), BTU/scf of waste gas. The heat content of the waste gas is a function of the VOC content. (The waste gas pressure is not a design consideration, because the majority of waste gases enter the incinerator at atmospheric pressure.)

Step 2. For thermal incinerators, determine the combustion temperature (T_5) based on the desired VOC destruction efficiency. Suggested combustion temperature (T_5) values for waste gases containing nonhalogenated VOCs are 1,600°F and 1,800°F, respectively, for 98 and 99 percent VOC destruction efficiencies. These temperatures correspond to a 0.75-second residence time in the incinerator. Higher temperatures of about 2,000°F (and 1-second residence time) are required for the destruction of halogenated VOCs by 98 percent or more.

For catalytic incinerators, select the preheat temperature (T_5) considering the amount and type of VOCs present in the waste gas. Upon oxidation in the catalyst bed, the heat content of the waste gas VOCs is released in the bed. Consequently, the temperature of the bed and waste gas increase, as explained in Section 3.1.1, by about 27°F per each one percent of VOC LEL. For most VOCs oxidized on precious metal catalysts, the suggested preheat temperature (T_5) is 600°F. For a waste gas content of 4 to 10 percent LEL, this results in an average catalyst operating temperature of 700 to 900°F (i.e., 600°F + [4 to 10]% LEL x 27°F/%LEL ≅ 700 to 900°F).

Less stable VOCs, such as monohydric alcohols, aromatic hydrocarbons or propylene, can use lower catalyst operating temperatures, on the order of 500°F.

In rare cases, where the VOC content in the waste gases is less than 2 percent LEL, higher preheat temperatures (700-750°F), or more catalyst than typically required, may have to be used to obtain equivalent 700° to 900°F catalyst operating temperature and VOC destruction. However, to avoid this increase in fuel/catalyst costs, the VOC content of the waste gas is increased when possible by repeated recycling of waste gases within the process being controlled, before introducing them to the incinerator.

Waste gases with VOC contents higher than 20 percent LEL are not suitable for catalytic incineration, because the VOC heat content released increases the catalyst bed temperature to beyond 1,200°F, the maximum permissible temperature to which the catalyst bed can be exposed continuously.

Conversely, for the destruction of more stable VOCs, such as methane, higher preheat temperatures (900-1,000°F) are used. This, however, requires a lower maximum allowable VOC content--i.e., 7 to 11%.

Step 3. (Applies to only catalytic incinerators.) Calculate the catalyst bed outlet temperature (T_6). As indicated in Step 2, T_6 should not exceed 1,200°F. Further, as explained in Section 3.1.1, the heat of combustion released by the VOC in the waste gas increases the waste gas temperature by 27°F for each one percent of LEL. Because the average heat capacity of air is approximately 0.018 BTU/scf-°F, the waste gas heat content required to increase the waste gas temperature by 27°F would be equivalent to 0.5 BTU/scf. (0.018 x 27). It follows that a waste gas VOC heat content of 1 BTU/scf would, when released through combustion, increase the waste gas temperature by about 55°F upon release (2 x 27°F). Therefore, we can write:

$$T_6 = T_5 + (Q_1/Q_5)(55)(h_1) \qquad (3\text{-}1)$$

where T_5 = Preheat temperature, °F, determined from Step 2

h_1 = Waste gas heat content, (BTU/scf of waste gas)

Q_1/Q_5 = Ratio of the waste gas flow rate at the incinerator inlet to the flue gas flow rate at the preheat chamber exit. For dilute waste gases consider the ratio to be equal to 1, by neglecting the increase in flue gas volume due to the fuel addition.

$T_6 \leq 1200$°F

Step 4. Determine the waste gas temperature at the incinerator inlet (T_2) = the temperature at the heat exchanger outlet. With thermal incineration, T_2 must be no higher than 1,000 to 1,100°F, to avoid preignition of the waste gas before it reaches the combustion chamber. T_2 is also limited by the heat VOC content of the inlet waste gas. As eq. 3-1 indicates, the temperature rise realized in the combustion chamber is proportional to this heat content. For example, if a 20 percent LEL stream were incinerated, the temperature rise would be about 540°F (20 x 27°). Subtracting this from a combustion temperature of 1500°F yields a maximum T_2 of about 950°F. For catalytic incinerators, the waste gas temperature (T_2) entering the preheat chamber can be as high as the preferred preheat temperature (T_5), 600°F.

Broadly speaking, as T_2 increases, the auxiliary fuel requirement (and cost) decreases. But at the same time, the size and cost of the recuperative heat exchanger increases, driving up the total capital investment of the system. Thus, there is a trade-off between capital and operating costs, the extent of which depends upon the value of T_2 selected. However, as this selection depends on the results of a process optimization analysis, no firm guidance can be given for selecting T_2--except that it should not exceed the design limits noted above. The discussion that follows may guide one in performing such an optimization.

First of all, the waste gas temperature entering the combustion chamber/preheat chamber is determined by the heat exchanger heat transfer performance (denoted by ϕ or HE) and the waste gas inlet temperature (T_1).

For incinerator systems with no heat exchangers

$T_2 = T_1$ (the temperature of the entering waste gas).

For incinerator systems with primary (recuperative) heat exchangers, T_2 is calculated from the known value of the system's heat exchanger heat transfer performance (ϕ) and the following expressions:

$T_2 = T_1 + \phi\ (T_5 - T_1)$ for thermal incinerators

$T_2 = T_1 + \phi\ (T_6 - T_1)$ for catalytic incinerators

(Note: Use of a secondary heat exchanger has no affect on the T_2 value.)

Theoretically, a heat exchanger heat transfer performance (ϕ) approaching 100 percent is possible. However, as the heat exchanger performance increases the required heat transfer area, the heat exchanger cost increases enormously, approaching infinity as ϕ approaches 100 percent. Therefore, heat exchanger ϕ is selected based on economic considerations, as well as the waste gas and flue gas data. Included among the waste gas and flue gas data which impose limitations on ϕ are: (1) the flue gas temperature at the heat exchanger outlet (T_7) in the case of either type of incinerator and (2) the waste gas VOC content in the case of thermal incinerators.

Depending upon the flue gas constituents, mainly moisture and corrosive elements, T_7 should be about 500 °F, due to possible equipment corrosion and condensation that may occur if it falls below this temperature.

The value of T_7 is related to ϕ by the following equation:

$$\phi = \frac{Q_5 Cp_5}{Q_2 Cp_2}\left[1 - \frac{T_7 - T_1}{T_5 - T_1}\right] \text{ for thermal incinerators} \qquad (3\text{-}4)$$

$$\phi = \frac{Q_6 Cp_6}{Q_2 Cp_2}\left[1 - \frac{T_7 - T_1}{T_6 - T_1}\right] \text{ for catalytic incinerator} \qquad (3\text{-}5)$$

where Q_1, Q_5, and Q_6 = waste gas flow rate and flue gas flow rates at the exit of combustion chamber and catalyst bed respectively (scfm)

Cp_2, Cp_5, and Cp_6 = Mean heat capacities of waste gas and of flue gas at the exit of the combustion chamber and catalyst bed, respectively (Btu/ft^3 °F)

It is evident from the above equation that maximum possible ϕ increases with the value of T_1. However, T_1 can never equal or exceed T_7.

The following heat transfer performance (ϕ) capabilities are commonly reported for typical modular heat exchangers: 35 to 40 percent for 1-pass, 45 to 50 percent for 2-pass, and 65 to 70 percent for 3-pass units.

Step 5. Calculate the amount of auxiliary fuel required (Q_3) based on the following energy balance around the combustion chamber/preheat chamber (see Figure 3-2):

Amount of sensible heat leaving	-	Amount of sensible heat entering	=	Amount of heat of VOCs released from combustion	+	Amount of heat of fuel released from combustion

$$(H_5 + H_L) - (H_2 + H_3) = Q_1\ h_1 + Q_3\ h_3 \text{ (for thermal incinerators)} \quad (3\text{-}6)$$

$$(H_5 + H_L) - (H_2 + H_3) = Q_3\ h_3 \quad (3\text{-}7)$$

(for catalytic incinerators, because VOC heat content is not released in the preheat chamber).

where H_5 = Sensible heat of flue gas at the combustion chamber/preheat chamber exit (BTU/min)
H_L = Heat loss = 10% H_5 (assumed)
H_2 = Sensible heat of waste gas at the incinerator/preheat chamber inlet (BTU/min)
H_3 = Sensible heat of fuel used (BTU/min)
= 0 (for fuel entering at ambient conditions)
Q_1 = Waste gas flow rate (scfm)
Q_3 = Auxiliary fuel flow rate (scfm)
h_1 = Waste gas heat content (BTU/scf of waste gas)
h_3 = Lower heating value (LHV) of fuel (BTU/scf).

By substituting $H_i = Q_i\ Cp_i\ \Delta T_i$, where Cp_i represents the mean heat capacity for the temperature difference of ΔT_i ($T_i - T_r$, the reference temperature) the sensible heat values of the gas streams leaving and entering the chamber can be expressed by the following equations:

$$H_5 = Q_5\ Cp_5\ \Delta T_5 \quad (3\text{-}8)$$

where Cp_5 = Mean heat capacity of the flue gas for the temperature interval of ΔT_5) from the reference temperature (70°F) to the combustion temperature (in the case of thermal incinerators) or to the preheat temperature (in the case of catalytic incinerators). Cp_5 = 0.0194, 0.0196, and 0.0198 BTU/ft^3°F for thermal incinerator combustion temperatures of 1,600, 1,800, 2,000°F, respectively, and 0.0183 BTU/ft^3°F for a catalytic incinerator preheat temperature of 600°F

$$H_2 = Q_2\ Cp_2\ \Delta T_2 \quad (3\text{-}9)$$

where C_{p2} = Mean heat capacity of the waste gas for the temperature interval ΔT_2 from the reference temperature (70°F) to the temperature at the inlet of the incinerator/preheat chamber (T_2). $C_{p2} = 0.0181$ BTU/ft^3°F for waste gas temperatures (T_1) up to 300°F

$Q_2 = Q_1$.

Substituting the above enthalpy values for H_2 and H_5 and $Q_1 + Q_3 = Q_2 + Q_3$ for Q_5 in eq. (3-6) or (3-7), rearranging the resulting equation, and accounting for the heat loss yields the following (see Appendix 3B for details):

$$\frac{Q_3}{Q_2} = \text{Fuel used, ft}^3\text{/std. ft}^3\text{ of waste gas}$$

$$= \frac{1.1\ C_{p5}\ \Delta T_5 - C_{p2}\ \Delta T_2 - h_1}{h_3 - 1.1\ C_5\ \Delta T_5} \qquad (3\text{-}10)$$

Step 6. Calculate the gas flow rate leaving the combustion chamber/preheat chamber (Q_5). In the case of a thermal incinerator, because the combustion takes place primarily in the combustion chamber, the resulting flue gas consists of the products of combustion of the waste gas VOCs and the auxiliary fuel. In the case of catalytic incinerators, because fuel is added in the preheating chamber to raise the waste gas temperature for subsequent combustion in the catalyst bed, the gases exiting the preheat chamber consist of the waste gas and the products of fuel combustion.

In the incineration of dilute VOC waste gases the flue gas flow rate is approximately equal to the total of the waste gas flow rate and fuel flow rate used for combustion, because no outside combustion air is used (i.e., $Q_4 = 0$). (This is because, in the case of thermal incinerators, the increase in flow rate due to combustion of the waste gas VOCs and fuel is negligible.) Therefore, as indicated above:

$$Q_5 = Q_2 + Q_3 \tag{3-11}$$

where Q_5 = Flue gas flow rate (scfm)
Q_2 = Waste gas flow rate (scfm)
Q_3 = Auxiliary fuel burned (scfm)

Item 2. Establish the Amount of Catalyst Required.

The cost of catalyst consumed or replaced represents a significant catalytic incinerator operating cost item. Therefore, the amount of catalyst used and its expected life must be accurately determined in the initial design of a catalytic incinerator system.

The amount of catalyst required usually depends upon the type and age of catalyst used, types and amounts of VOCs encountered, the destruction efficiency required, and the amounts of potential reactants present that impair the catalyst activity, which is usually highest when the catalyst is fresh. Because in most cases only limited information is available about the potential reactants and noncombustible particulate matter that may be present in the waste gas, most catalytic incinerator systems are overdesigned to compensate for the unexpected. The amounts of catalyst required for the destruction of various types of VOCs are usually established by laboratory tests by catalyst manufacturers.

The amounts of precious metal catalysts commonly used are 1.5 and 2 ft^3 per 1,000 scfm waste gas (equivalent to space velocities of 40,000 and 30,000 hr^{-1}) for 90 and 95 percent VOC destruction efficiencies, respectively.(1) The corresponding amounts of base metal catalysts used are 4 to 6 ft^3/1,000 scfm (i.e., space velocities of 15,000 to 10,000 hr^{-1}).(2)

Item 3. Determine the Pressure Drop Required Across the System.

The total pressure drop required across an incinerator system determines the waste gas fan size and horsepower requirements, which, in turn, determine the fan capital cost and electricity consumption.

The total pressure drop across an incinerator system depends on the number and types of equipment included in the system and on design considerations. The estimation of actual pressure drop requirements involves complex calculations based on the specific system's waste gas and flue gas conditions and equipment used. For the purposes of this section, however, the following approximate values can be used:

Pressure drop (ΔP, in. H_2O) across

Thermal incinerators = 4
Catalytic incinerators = 6
Heat exchangers: 35% = 4
50% = 8
70% = 15

Once the total pressure drop required is estimated (as a summation of the pressure drops across all pieces of equipment in the incinerator system), the blower electricity requirements can be estimated from the basic fan horsepower requirement equation. For example, at a combined fan-motor efficiency of 62 percent, the fan horsepower equation dictates that 0.19 kWh of electricity per hour is required for moving 1,000 acfm of flow rate at a 1-in. water pressure drop. Therefore, the fan power requirement is estimated by multiplying the total pressure drop (inches of water) by the total flue gas flow rate (thousand acfm) and the 0.19 kWh/hr factor. For the cases where an additional fan is used to add outside air to the system, its power requirement must be calculated separately. (See Section 2 of this Manual for more on how to estimate electricity requirements.)

3.3 Estimating Total Capital Investment

This section provides a general methodology for developing "study" estimates of capital costs (April 1986 dollars) for thermal and catalytic incinerator systems. The precision of the "study" cost estimate (i.e., ±30 percent) applies to the estimates presented herein. In this method, the total purchased cost of the system equipment is the basis to which predetermined factors are applied, to estimate the system direct and indirect installation costs. (See Section 2 for a more thorough discussion of this estimating method.)

The purchased cost of an incinerator varies widely depending upon several design factors. Therefore, discretion is needed when using generic incinerator cost information. Among the factors that influence the incinerator purchased cost are supplier's design experience, materials of construction, instrumentation, the type of heat exchanger used and the nature of the installation, i.e., whether indoor, outdoor, ground level, or roof top.

The nature of the installation has a particular effect on the system design and, consequently, the cost. Specifically, incinerator systems for roof top installations are made of light weight material, while equipment for outdoor locations can be preassembled in larger modules than for indoor locations. Traditionally, for thermal incinerators stainless steel combustion chambers are used to achieve a maximum equipment life (typically 15 years). Several manufacturers offer carbon steel units at costs of 60 to 70 percent of that for stainless steel units. Of course, the life of such equipment will be lower--approximately 10 years. Catalytic incinerators will tolerate lower grade material (e.g., carbon steel) better than thermal units because of their lower operating temperatures. However, catalyst bed enclosures are mostly fabricated of stainless steel.

3.3.1 Thermal Incinerator Equipment Costs

Figure 3-3 presents thermal incinerator equipment costs (April 1986 dollars) as a function of the waste gas volume flow rate at standard conditions of 70°F and 0 psig. This figure was developed from cost information received from three incinerator manufacturers for three volume flow rates each.(3,4,5) Analytical equations for these equipment cost curves are presented in Table 3-1. (The table also presents analytical equations for catalytic incinerators, the cost curves for which are discussed in Section 3.3.2.) The equipment costs listed represent all the equipment in an incinerator system including a combustion chamber with burner, waste gas fan, inlet and outlet plenums, prepiping and prewiring, instrumentation and controls, a 10-ft stack, and in the case of heat recovery, a primary heat exchanger. The cost data apply to the cases of dilute VOC content waste gases (to which no outside air is added) containing up to 25 percent LEL VOC and incinerated at a 1500°F combustion temperature.

The Figure 3-3 (or Table 3-1) cost data can be applied to other cases of waste gas compositions including those requiring slightly different combustion temperatures without introducing significant errors to the costs. When these data are applied to different combustion temperatures,

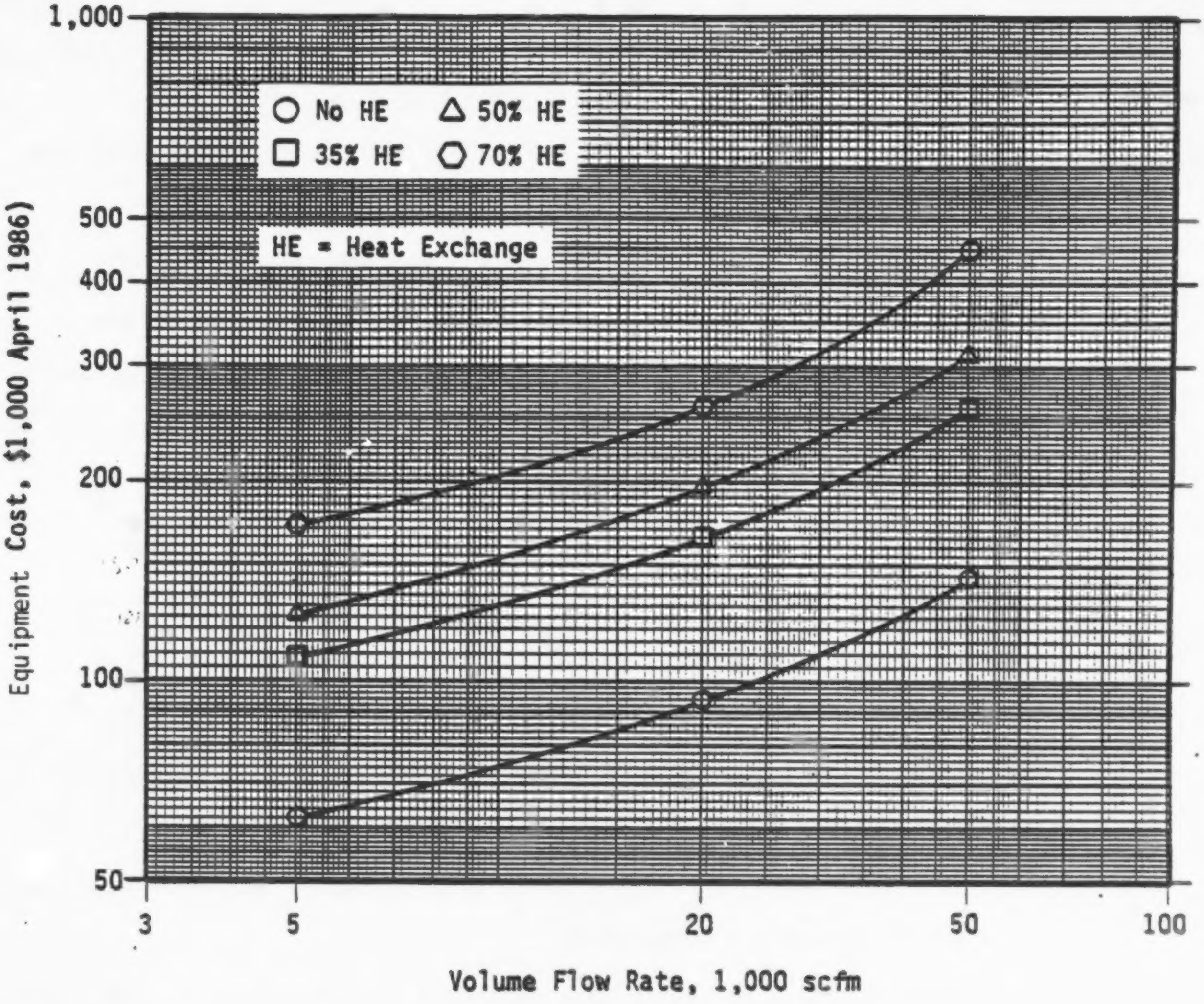


Figure 3-3. Thermal incinerator equipment cost estimates

TABLE 3-1. EQUATIONS FOR INCINERATOR EQUIPMENT COSTS[a] (APRIL 1986 DOLLARS)

Extent of heat exchange	Equipment cost equation for thermal incinerators[b]	Equipment cost equation for catalytic incinerators
None	$\ln Eq = [14{,}402 - 992(\ln Q) + 70\,(\ln Q)^2]\,(10^{-3})$	$\ln Eq = [24{,}086 - 3{,}252(\ln Q) + 205\,(\ln Q)^2](10^{-3}$
35%	$\ln Eq = [16{,}175 - 1{,}262(\ln Q) + 85\,(\ln Q)^2]\,(10^{-3})$	$\ln Eq = [27{,}170 - 3{,}789(\ln Q) + 231\,(\ln Q)^2](10^{-3}$
50%	$\ln Eq = [15{,}784 - 1{,}165(\ln Q) + 81\,(\ln Q)^2]\,(10^{-3})$	$\ln Eq = [26{,}497 - 3{,}650(\ln Q) + 225\,(\ln Q)^2](10^{-3}$
70%	$\ln Eq = [20{,}608 - 2{,}119(\ln Q) + 131\,(\ln Q)^2](10^{-3})$	$\ln Eq = [21{,}685 - 2{,}643(\ln Q) + 174\,(\ln Q)^2](10^{-3}$

[a] These equations should not be extrapolated outside the flowrate range of 5,000 to 50,000 scfm.

[b] Eq - Equipment cost in April 1986 dollars, Q - Volume flow rate in scfm.

it is assumed that the system will tolerate minor adjustments to the incinerator face velocity and residence time values resulting from these temperatures. Even if the system size were adjusted to compensate for these different temperatures,the resulting changes in the equipment cost would be minor. For example, if a combustion chamber were sized for a combustion temperature of 1,700°F instead of 1,500°F the cost increase due to the increase in combustion chamber size would be less than 5 percent.

The Figure 3-3 and Table 3-1 cost data can also be applied to cases of waste gas compositions to which outside ambient air is added for combustion. In these cases, the flue gas flow rate at standard conditions must be substituted for the waste gas flow rate in the figure.

3.3.2 Catalytic Incinerator Equipment Costs

Figure 3-4 presents catalytic incinerator equipment costs (April 1986 dollars) as a function of the waste gas volume flow rate at standard conditions (70°F and 0 psig), developed from cost information received from five equipment manufacturers.(6-10) Analytical equations representing these cost data are presented in Table 3-1. The equipment cost data represent all the equipment in an incinerator system, including the burner, fan, housing, skid mounting, instrumentation and controls, a 10-ft stack, catalyst, and where heat recovery is used, a primary heat exchanger. The cost data apply to dilute VOC waste gases requiring a temperature of 600°F at the preheat chamber exit/catalytic bed inlet. The cost data can be applied to waste gases to which outside air is added by substituting the flue gas flow rate at standard conditions for the waste gas flow rate in Figure 3-4.

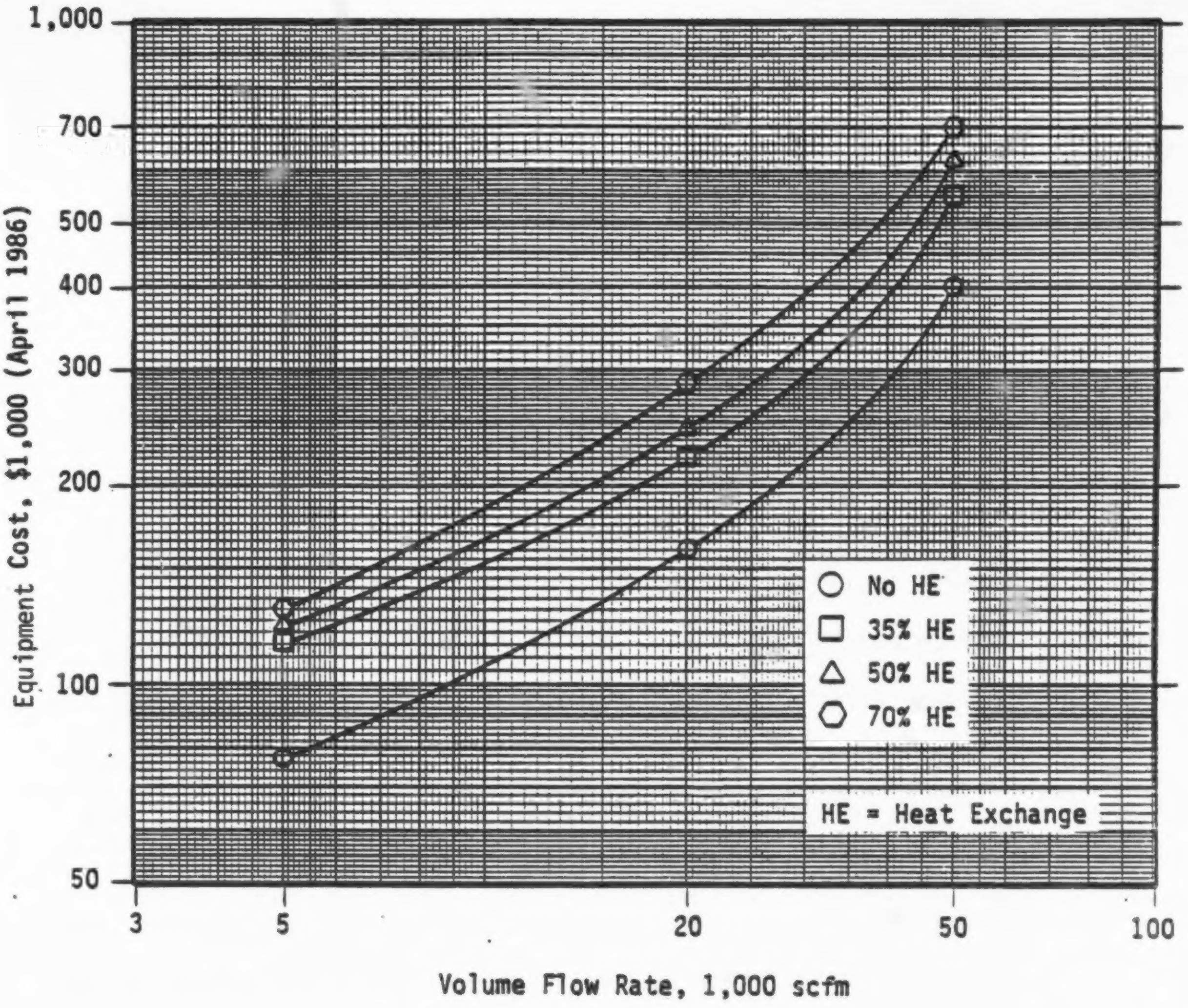


Figure 3-4. Catalytic incinerator equipment cost estimates

3.3.3 Total Capital Investment for Incinerators

The total incinerator capital investment as a percentage of equipment cost varies with the type of installation, i.e., whether custom or package. The type of incinerator installation is an important consideration. For most nonchemical industry operations, such as drying ovens, painting facilities, printing and coating operations and other processes where incinerators are commonly used for VOC control, waste gas flow rates generated are small. Modular incinerator units, prefabricated and preassembled, are most suitable for such flow rates. The size of skid-mounted modular units is limited to handling flows up to about 20,000 cfm, mainly because of limitations to the sizes of trucks carrying them and access doors at the plant sites. Of course, larger units, prefabricated and partially assembled, are possible for outdoor installations. Some incinerators handling up to 100,000 cfm, which are partly field-fabricated and field-assembled, can be found in chemical plant applications. Typically, heat exchangers can accommodate gas flows from 3,000 to 20,000 cfm. For larger flowrates, multiple heat exchangers units are usually installed in parallel in a single incinerator system.

For incinerators, the total capital investment varies from a small percentage (about 110 percent) to more than 200 percent of the total equipment purchased cost. The custom installation charges of an incinerator system, including incinerator and heat exchanger units, ductwork, fans, and a tall stack if required, may amount to 300 to 400 percent of the total equipment purchased cost, depending upon the installation.

The skid-mounted units ready to be placed on a concrete pad at a prepared site incur the lowest installation expenses. Roof top installations will require higher costs, field-assembled units even higher, and field-fabricated and assembled units the highest costs.

For estimating purposes, the total capital investment (TCI) of a skid-mounted modular unit can be calculated at 125 percent of the total purchased equipment cost. Conversely, Table 3-2 presents typical estimated capital cost factors applicable to custom installations of both thermal and catalytic incinerator systems. No cost factor is included for estimating ductwork cost in the table because the ductwork size and length and, consequently, its cost will vary with the installation and the waste gas flowrate. For both packaged and custom incinerator units, the ductwork capital cost must be estimated separately. Alternatively, in the case of custom units, the purchased cost of the ductwork and other auxiliary equipment can be added to the total incinerator equipment cost prior to applying the Table 3-2 factors for estimating the total capital investment. (Note: As Section 2 of the *Manual* indicates, the TCI also includes costs for land, working capital, and off-site facilities, which are not included in the direct/indirect installation factor. However, as these items are rarely required with incinerator systems, they will not be considered here. Further, no factor has been provided for site preparation (S.P.), or buildings (Bldg.) since these site-specific costs depend very little on the purchased equipment cost. Lastly, for the incinerator cases in which outside air is added for combustion or dilution, the cost of the fan used for this purpose also must be estimated separately.)

3.4 Estimating Total Annual Cost

3.4.1 Direct Annual Costs

For incinerator systems with no heat exchanger, the cost of auxiliary fuel is the major direct annual cost. Fuel costs are considerably

TABLE 3-2. CAPITAL COST FACTORS FOR CUSTOM THERMAL AND CATALYTIC INCINERATORS[a]

Cost Item	Cost Factor (Fraction of Indicated Cost)
DIRECT COSTS	
1) Purchased equipment cost	
Incinerator	As required
Auxiliary equipment[b]	As required } A
Instruments and controls[c]	0.1 A
Taxes	0.03 A
Freight	0.05 A
Total Purchased Equipment cost	B(=1.18A)
2) Direct installation costs	
Foundations and supports	0.08 B
Erection and handling	0.14 B
Electrical	0.04 B
Piping	0.02 B
Insulation	0.01 B
Painting	0.01 B
Site preparation (S.P) Buildings (Bldg.)	As required
Total Installation Direct Costs	0.30 B + S.P. + Bldg.
Total Direct Cost	1.30 B + S.P. + Bldg.
INDIRECT COSTS	
Engineering and supervision	0.10 B
Construction and field expenses	0.05 B
Construction fee	0.10 B
Start-up	0.02 B
Performance test	0.01 B
Contingency	0.03 B
Total Indirect Costs	0.31 B
Total Direct and Indirect costs = Total Capital Investment	1.61 B + S.P. + Bldg.

[a] Reference 11.

[b] Includes ductwork and any other equipment normally not included with unit furnished by incinerator vendor.

[c] Cost of instrumentation and controls often furnished by vendor.

lower for catalytic incinerators than for thermal units because of the lower operating temperature. Besides the fuel, the utility requirements for an incinerator include electricity for the waste gas fan and, if used, the ambient air fan. Operating labor is usually small for both types of incinerators--0.5 hr/shift.(11)

Maintenance costs of well-designed and maintained thermal or catalytic incinerators of either type are low, with a substantial portion required for the heat exchanger maintenance. (Well-designed and fabricated heat exchangers, i.e., stainless heat exchangers, usually have lower maintenance expenses.) The incinerator unit maintenance requirements may include repairs to the refractory lining and blowers, maintenance of control instruments, and cleaning of flame rods, if used. Maintenance labor may be estimated at 0.5 hr/shift. The maintenance materials may be assumed to equal the maintenance labor cost.(11)

Depending upon the incinerator and how it is operated, the life of a given load of catalyst may be 2 to 10 years. A conservative estimate of catalyst replacement cost can be based on the lower life time--2 years. The initial costs of precious metal and base metal (manganese dioxide) catalysts are $\$3,000/ft^3$(1) and $\$600/ft^3$ (2), respectively. Item 2 of Section 3.2.2 showed how to estimate the catalyst requirement (ft^3). Finally, the catalyst replacement labor is minimal compared to the catalyst cost.

Based on these values and the estimating procedure shown in Section 2, the catalyst replacement cost (CRC_{cat}) would be:

$$CRC_{cat} = C_{cat} \times 1.08 \times CRF_{cat} \qquad (3\text{-}12)$$

TABLE 3-3. SUGGESTED FACTORS FOR ESTIMATING INCINERATOR ANNUAL COSTS

Item	Suggested Factor
Direct Operating Costs	
Operating labor[a]	0.5 hr/shift
Supervisory labor[a]	15% of operating labor
Maintenance labor	0.5 hr/shift
Maintenance materials[a]	100% of maintenance labor
Replacement parts	Thermal incinerators: None Catalytic incinerators: (See eq. 3-12)
Utilities:	
Fuel	The amount of fuel required is calculated from Step 5 (Q_3) of Section 3.2.
Electricity[b,c]	Use the following Δ P values in estimating electricity requirements: Thermal incinerators = 4 in. water Catalytic incinerators = 6 in. water Heat exchange of 35% = 4 in. water 50% = 8 in. water 70% = 15 in. water Ductwork and stack = As required
Indirect Operating Costs	
Overhead	60% of sum of operating, supervisory, and maintenance labor and maintenance materials
Administrative charges	2% X TCI[d]
Property tax	1% X TCI[d]
Insurance	1% X TCI[d]
Capital recovery cost[e]	CRFs X [TCI - 1.08 x C_{cat}]

[a] Reference 11.

[b] The total Δ P of an incinerator system is the sum of base (i.e., incinerator) Δ P + heat exchanger Δ P.

[c] An equation to calculate electricity requirements (kWh per hour) is given in Item 3 of Section 3.2.2.

[d] TCI = Total capital investment.

[e] The CRFs (system capital recovery factor) is a function of the equipment life (10 years, typically) and the opportunity cost of the capital (i.e., interest rate). For instance, for a 10-year life and a 10% interest rate, $CRF_s = 0.1628$.

where: C_{cat} = initial catalyst cost ($)

1.08 = factor accounting for taxes and freight

CRF_{cat} = capital recovery factor for catalyst (e.g., for a 2-year life and a 10% interest rate, the factor would be 0.5762).

3.4.2 Indirect Annual Costs

As Table 3-3 shows, the indirect (fixed) annual operating costs include capital recovery; overhead; and property taxes, insurance, and administrative charges. The last three costs can be estimated at 1 percent, 1 percent, and 2 percent of the total capital investment, respectively. The system capital recovery cost is based on an estimated 10-year equipment life. (The incinerator system life varies from 5 to 15 years depending on the material of construction used and the durability of the design.) However, as Section 2 indicates, the system capital recovery cost is the product of the system capital recovery factor (CRF_s) and the total capital investment (TCI) less the purchased cost of the catalyst ("C_{cat} x 1.08"), This offset is necessary to avoid double-counting.

Finally, overhead may be estimated at 60% of the sum of operating, supervisory, and maintenance labor and maintenance materials. (See Section 2 of the *Manual* for a detailed discussion of the items comprising the total annual cost.)

References for Section 3

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2. Telephone conversation: Bob Campbell (Pillar Technologies, Inc., Hartland, WI) with Tom Rhoads (Pacific Environmental Services, Inc., Durham, NC), February 28, 1986.

3. Correspondence: Ralph N. Stettenbenz (C-E Air Preheater, Combustion Engineering, Inc., Wellsville, NY) to Vishnu S. Katari, April 3, 1986.

4. Correspondence: Ronald J. Baschiere (Hirt Combustion Engineers, Montebello, CA) to Vishnu S. Katari, April 7, 1986.

5. Correspondence: C.L. Bumford (Peabody Engineering, Stamford, CT) to Michael K. Sink (Pacific Environmental Services, Inc., Durham, NC), March 24, 1986.

6. Correspondence: Robert M. Yarrington (Engelhard Corporation, Union, NJ) to Vishnu S. Katari, April 30, 1986.

7. Correspondence: Kent Lewis (C-E Air Preheater, Combustion Engineering, Inc., Wellsville, NY) and Vishnu S. Katari, June 16, 1986.

8. Correspondence: Robert Hablewitze (Pillar Technologies, Inc., Hartland, WI) to Vishnu S. Katari, March 3, 1986.

9. Correspondence: Thomas Schmidt (Energy Development Associates, Itascon, IL) to Michael K. Sink, March 7, 1986.

10. Correspondence: Richard J. Ihde (TEC Systems, DePere, WI) to Michael K. Sink, March 11, 1986.

11. Vatavuk, William M. and Neveril, Robert. "Estimating Costs of Air-Pollution Control Systems, Part II: Factors for Estimating Capital and Operating Costs." Chemical Engineering, November 3, 1980, pp. 157-162.

APPENDIX 3A. LOWER AND UPPER EXPLOSIVE LIMITS FOR VOCs

TABLE 3A-1. FLAMMABILITY CHARACTERISTICS OF COMBUSTIBLE ORGANIC COMPOUNDS IN AIR[a,b]

	Mol. wt.	LEL,[c] % vol.	UEL,[d] % vol.
Methane	16.04	5.0	15.0
Ethane	30.07	3.0	12.4
Propane	44.09	2.1	9.5
n-Butane	58.12	1.8	8.4
n-Pentane	72.15	1.4	7.8
n-Hexane	86.17	1.2	7.4
n-Heptane	100.20	1.05	6.7
n-Octane	114.23	0.95	3.2
n-Nonane	128.25	0.85	2.9
n-Decane	142.28	0.75	5.6
n-Undecane	156.30	0.68	
n-Dodecane	170.33	0.60	
n-Tridecane	184.36	0.55	
n-Tetradecane	108.38	0.50	
n-Pentadecane	212.41	0.46	
n-Hexadecane	226.44	0.43	
Ethylene	28.05	2.7	36
Propylene	42.08	2.4	11
Butene-1	56.10	1.7	9.7
cis-Butene-2	56.10	1.8	9.7
Isobutylene	56.10	1.8	9.6
3-Methyl-Butene-1	70.13	1.5	9.1
Propadiene	40.06	2.6	
1,3-Butadiene	54.09	2.0	12
Acetylene		2.5	100
Methylacetylene		1.7	
Benzene	78.11	1.3	7.0
Toluene	92.13	1.2	7.1
Ethylbenzene	106.16	1.0	6.7
o-Xylene	106.16	1.1	6.4
m-Xylene	106.16	1.1	6.4
p-Xylene	106.16	1.1	6.6
Cumene	120.19	0.88	6.5
p-Cymene	134.21	0.85	6.5
Cyclopropane	42.08	2.4	10.4
Cyclobutane	56.10	1.8	
Cyclopentane	70.13	1.5	
Cyclohexane	84.16	1.3	7.8
Ethylcyclobutane	84.16	1.2	7.7

TABLE 3A-1. FLAMMABILITY CHARACTERISTICS OF COMBUSTIBLE ORGANIC COMPOUNDS IN AIR[a,b] (concluded)

	Mol. wt.	LEL, % vol.	UEL, % vol.
Cycloheptane	98.18	1.1	6.7
Methylcyclohexane	98.18	1.1	6.7
Ethylcyclopentane	98.18	1.1	6.7
Ethylcyclohexane	112.21	0.95	6.6
Methyl alcohol	32.04	6.7	36
Ethyl alcohol	46.07	3.3	19
n-Propyl alcohol	60.09	2.2	14
n-Butyl alcohol	74.12	1.7	12
n-Amyl alcohol	88.15	1.4	10
n-Hexyl alcohol	102.17	1.2	7.9
Dimethyl ether	46.07	3.4	27
Diethyl ether	74.12	1.9	36
Ethyl propyl ether	88.15	1.7	9
Diisopropyl ether	102.17	1.4	7.9
Acetaldehyde	44.05	4.0	36
Propionaldehyde	58.08	2.9	14
Acetone	58.08	2.6	13
Methyl ethyl ketone	72.10	1.9	10
Methyl propyl ketone	86.13	1.6	8.2
Diethyl ketone	86.13	1.6	
Methyl butyl ketone	100.16	1.4	8.0

[a] Reference 1.

[b] Most common handbooks (e.g., Reference 2) provide flammability information for VOCs.

[c] LEL - lower explosive limit

[d] UEL - upper explosive limit

TABLE 3A-2. CATALYTIC IGNITION TEMPERATURE FOR 90% CONVERSION[a]

Component	Temperature, °F
Hydrogen	220
Acetylene	395
Carbon monoxide	425
Propyne	460
Propadiene	480
Propylene	500
Ethylene	550
n-Heptane	575
Benzene	575
Toluene	575
Xylene	575
Ethanol	600
Methyl ethyl ketone	700
Methyl isobutyl ketone	700
Propane	770
Ethyl acetate	775
Dimethyl formamide	800
Ethane	810
Cyclopropane	850
Methane	920

[a] Reference 3.

References for Appendix 3A

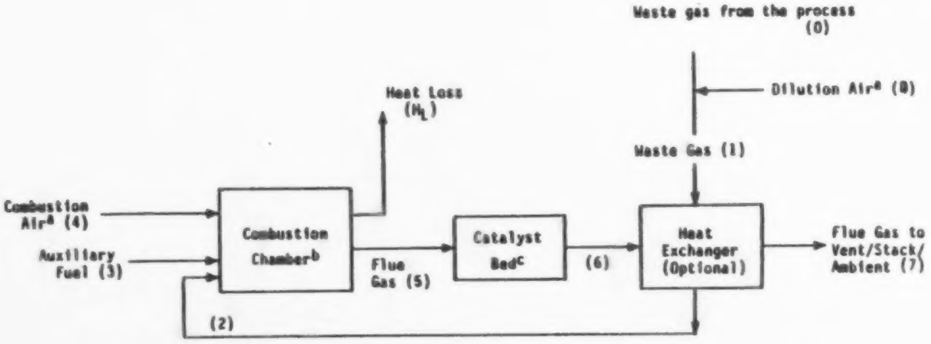
1. Hilado, Carlos J., "How to Predict if Materials Burn," Chemical Engineering, December 14, 1970, pp. 174-178.

2. Handbook of Chemistry and Physics, 54th Edition. Cleveland: The Chemical Rubber Co., 1973-1974, pp. D85-D92.

3. Organic Chemical Manufacturing Volume 4: Combustion Control Devices, (EPA-450/3-80-026). Research Triangle Park: U.S. Environmental Protection Agency, Emissions Standards and Engineering Division. December 1980. Report 3, p. I-2.

APPENDIX 3B

This appendix presents equations to estimate flue gas and fuel requirements of incineration systems combusting all types of waste gas compositions. These equations have been derived from energy and material balances around the incinerator system. Figure 3B-1, a reproduction of Figure 3-2, Section 3.2, is a schematic of an incinerator system. (All streams shown in Figure 3B-1 are also identified by the same subscripts in the energy and material balances)


Waste gas from the process (0)
Dilution Air^a (8)
Heat Loss (H_L)
Waste Gas (1)
Combustion Air^a (4)
Auxiliary Fuel (3)
Combustion Chamber^b
Flue Gas (5)
Catalyst Bed^c
(6)
Heat Exchanger (Optional)
Flue Gas to Vent/Stack/Ambient (7)
(2)

(a) When required.

(b) Referred to as Preheat Chamber in the case of catalytic incinerators.

(c) Included only in catalytic incinerators.

Figure 3B-1. Schematic diagram of an incinerator system.

Energy Balance

The basic energy balance equation around the combustion chamber is:

$$\begin{bmatrix}\text{Sensible heat}\\ \text{leaving}\end{bmatrix} - \begin{bmatrix}\text{Sensible heat}\\ \text{entering}\end{bmatrix} = \begin{bmatrix}\text{Heat released}\\ \text{from the}\\ \text{combustion of}\\ \text{VOCs}\end{bmatrix} + \begin{bmatrix}\text{Heat released}\\ \text{from the combustion}\\ \text{of fuel}\end{bmatrix}$$

From Figure 3B-1, $(H_5 + H_L) - (H_2 + H_3 + H_4) = Q_1h_1 + Q_3h_3$

where:

H = Sensible heat of the stream identified by the subscript, Btu/min

Q = Flow rate of the stream identified by the subscript, scfm

h_1 = Waste gas heat content released upon combustion, BTU/scf of waste gas
h_3 = Lower heating value (LHV) of fuel (BTU/scf)

Assume H_3 = 0 for fuel at ambient temperature
H_4 = 0 for ambient air
H_L = 10% of H_5

Therefore, $1.1\ H_5 - H_2 = Q_1\ h_1 + Q_3\ h_3$

Substituting the sensible heat expression $H = Q\ Cp\ \Delta T$ in the above equation yields:

$$1.1\ Q_5\ Cp_5\ \Delta T_5 - Q_2\ Cp_2\ \Delta\ T_2 = Q_1\ h_1 + Q_3\ h_3 \qquad (3B\text{-}1)$$

where Q = Flow rate (scfm)
C_p = Mean heat capacity for the temperature interval of ΔT, BTU/ft^3 °F
ΔT = Temperature interval from T_R (70°F) to T identified by the subscript
h_1 = Waste gas heat content (BTU/scf of waste gas)
h_3 = Lower heating value (LHV) of natural gas BTU/ft^3

Material Balance (see Figure 3B-1)

An (approximate) material balance around the combustion chamber yields:*

$$Q_5 = Q_2 + Q_3 + Q_4 \qquad (3B\text{-}2)$$

where Q_5 = Flue gas flow rate (scfm)
Q_2 = Waste gas flow rate entering the system (scfm)
Q_3 = Auxiliary fuel burned (scfm)
Q_4 = Ambient air used for combustion (scfm)

The quantities of Q_2, Q_3, and Q_4 are estimated as follows:

1) Waste gas entering the combustion chamber (Q_2)

= Q_0 For cases of waste gases to which no dilution air is added (i.e., Cases 1, 2, and 4 through 6 in Table 3B-1)

= Q_0 (he/hd) For cases of waste gases to which dilution air is added (i.e., Case 3 in Table 3B-1) (3B-3)

where h_e = Waste gas heat content before dilution, BTU/scf
h_d = Waste gas heat content after dilution.

*Strictly speaking, equation 3B-2 is a mole balance, not a material balance. However, because the moles (or volumes) of the reactants (VOC, fuel, air, etc.) entering the combustion chamber/catalyst bed approximately equal the moles of products leaving (H_2O, CO_2, N_2, etc.), equation 3B-2 is accurate--at least for purposes of this Manual.

2) Ambient air used (Q_4)

Q_4 = 0 For dilute VOC waste gases (category 1 in Table 3B-1)

$$Q_4 = Q_{4,VOC} + Q_{4,Fuel} \qquad (3B\text{-}4)$$

where $Q_{4,VOC}$ = Ambient air used for the combustion of the VOC

$= Q_2 \, [\Sigma m_i \, (X + Y/4 - Z/2)_i - m_{O_2}] \, 4.79$

where m_i = Volume content of ith VOC component in the waste gas

X, Y, Z = Atoms of C, H_2, and O_2 in the VOC

m_{O_2} = Volume content of oxygen in the waste gas

$Q_{4,Fuel}$ = Ambient air used for the combustion of the fuel

= 9.58 (base amount of fuel burned)(1 + E)

where E = Excess air used, % of total stoichiometric air requirement for fuel only

3) Auxiliary fuel burned (Q_3)

Q_3 = Base amount of fuel burned (calculated at zero excess air) + additional amount of fuel burned to ensure raising the excess air to the combustion temperature, (i.e., 9.58E [base amount of fuel])

= Base amount (1 + K)

where K = Excess amount used, % of base amount

The value of K can be determined as follows:

(Additional fuel used)

$= [9.58 \text{ (base amount of fuel) } E + \text{Additional fuel}] \, Cp_5 \, \Delta T_5 \div h_3$*

Therefore,

$$K = \frac{\text{Additional fuel used}}{\text{Base amount of fuel}} \qquad (3B\text{-}5)$$

$= 9.58E/[(h_3/Cp_5 \, \Delta T_5) - 1]$

= Constant for a given fuel and known values of E and T_5

= 0 for dilute VOC waste gases.

---

*In other words, the "additional fuel" must heat both the excess air used in combustion of the base fuel and itself, from the ambient (reference) temperature to the combustion temperature (T_5).

Estimation of Combustion Flue Gases Generated

Substituting the values of Q_2, Q_3, and Q_4 into the basic material balance equation (Equation 3B-2) yields the following expression:

$$Q_5 = Q_2 + Q_3 + Q_2 \left[\sum m_i \, (X + Y/4 - Z/2)_i - m_{O_2}\right] 4.79 + 9.58 \, (\text{base amount of fuel})(1 + E) \qquad (3B\text{-}6)$$

However, the base amount of fuel = $Q_3/(1 + K)$, so that:

$$Q_5 = Q_2 + Q_3 + Q_2 \left[\sum m_i \, (X + Y/4 - Z/2)_i - m_{O_2}\right] 4.79 + 9.58 \, Q_3 \, (1 + E)/(1 + K) \qquad (3B\text{-}7)$$

$$= Q_2 + Q_3 \text{ (for dilute VOC waste gases)}$$

The above equation applies to the waste gases to which ambient air is added for combustion purposes. However, in many cases, the waste gas contains air sufficient to provide partial or complete combustion air requirements. Table 3B-1 identifies various waste gas types and their ambient air and auxiliary fuel requirements. Table 3B-2 presents the equations for estimating the flue gas (Q_5) flow rate from the combustion of each type of waste gas.

Estimation of Actual Amount of Auxiliary Fuel Burned (Q_3)

Substituting the material balance equation (Equation 3B-2) into the energy balance equation (Equation 3B-1) and solving for Q_3/Q_2 (i.e., the amount of fuel required per unit of waste gas flow rate) yields the following expression:

$$\frac{Q_3}{Q_2} = \frac{1.1 \left[1 + (\Sigma m_i \, (X + Y/4 - Z/2)_i - m_{O_2}) 4.79\right] Cp_5 \Delta T_5 - \left[Cp_2 \Delta T_2 + h_1\right]}{(h_3 - 1.1 \, Cp_5 \, \Delta T_5) - 10.54 \, (1 + E) \, Cp_5 \, \Delta T_5/(1 + K)} \qquad (3B\text{-}8)$$

$$= \frac{1.1 \, Cp_5 \, \Delta T_5 - Cp_2 \, \Delta T_2 - h_1}{h_3 - 1.1 \, Cp_5 \, \Delta T_5} \text{ (for dilute waste gases)}$$

The values of all items in the above equation are specific to the waste gas composition and fuel type. Therefore, the equation can be further

TABLE 3B-1. CATEGORIZATION OF WASTE GAS STREAMS

Waste gas		
Category	Composition	Auxiliaries and other requirements
1	Mixture of VOC, air, and inert gas with >16% O_2 and a VOC content <25% LEL (i.e., heat content <13 Btu/ft^3)	Auxiliary fuel is required. No auxiliary air is required.
2	Mixture of VOC, air, and inert gas with >16% O_2 and a VOC content between 25 and 50% LEL (i.e., heat content between 13 to 26 Btu/ft^3)	Dilution air is required to lower the heat content to <13 Btu/ft^3. (Alternative to dilution air is installation of LEL monitors.)
3	Mixture of VOC, air, and inert gas with <16% O_2	Treat this waste stream the same as categories 1 and 2, except augment the portions of the waste gas used for fuel burning with outside air to bring its O_2 content to above 16 percent.
4	Mixture of VOC and inert gas with zero to negligible amount of O_2 (air) and <100 Btu/scf heat content	Oxidize it directly with a sufficient amount of air.
5	Mixture of VOC and inert gas with zero to negligible amount of O_2 (air) and >100 Btu/scf heat	Premix and use it as a fuel.
6	Mixture of VOC and inert gas with zero to negligible amount of O_2 and heat content insufficient to raise the waste gas to the combustion temperature	Auxiliary fuel and combustion air for both the waste gas VOC and fuel are required.

TABLE 3B-2. EQUATIONS FOR ESTIMATING FLUE GAS QUANTITIES

Waste gas category	Equations for flue gas flow rate (Q_5), scfm
1	$Q_5 = Q_2 + Q_3$, where $Q_2 = Q_0$
2	$Q_5 = Q_2 + Q_3$, where $Q_2 = Q_0\ (he/hd)$
3	$Q_5 = Q_2 + Q_3 + Q_{4,Fuel}$, where $Q_{4,Fuel} = 9.58\ (1+E)$ (Base amount of fuel used) $= 9.58\ Q_3\ (1+E)/(1+K)$
4	$Q_5 = Q_2 + Q_2\ [\Sigma m_i\ (X + Y/4 - Z/2)i - m_{O_2}]\ 4.79\ (1+E)$ Assumes no auxiliary fuel except for pilot is needed
5	Premix and use it as a fuel.
6	$Q_5 = Q_2 + Q_3 + 9.58\ Q_3\ (1+E)/(1+K) + Q_2\ [\Sigma m_i\ (X + Y/4 - Z/2)_i - m_{O_2}]\ 4.79$

simplified by substituting known values of these items.

The values of C_p and K are established based on the waste gas temperature, flue gas combustion temperature, and the type of fuel used. The overall C_p value of the flue gas is calculated more accurately from the C_p values of the flue gas components, which include CO_2, H_2O, O_2, and N_2. Alternatively, as an approximation, the C_p value of air can be substituted for C_p values of the flue gas and the waste gas. The mean heat capacity values for air are 0.0194, 0.0196, and 0.0198 BTU/ft^3°F for combustion temperatures of 1,600, 1,800, and 2,000°F, respectively. The mean heat capacity value of air, to use with a waste gas at 100°F, is 0.018 BTU/ft^3°F.

At a combustion temperature of 1,600°F, K = 0.033 for a LHV of 900 BTU/ft^3 for natural gas, an excess air of 10 percent, and a reference temperature of 70°F. When this value of K and the values of C_ps' and ΔTs' are substituted into equation 3B-8, assuming the waste gas temperature to be 100°F, we obtain:

$$\frac{Q_3}{Q_2} = \frac{1}{1{,}000}\left[60.2 - 1.8\ h_1 + 293\left(\Sigma\ m_i\ (X + Y/4 - Z/2)_i - m_{O_2}\right)\right] \quad (3B\text{-}9)$$

Table 3B-3 presents equations for estimating fuel requirements of all types of waste gases.

TABLE 3B-3. EQUATIONS FOR ESTIMATING FUEL REQUIREMENTS

Waste gas category	General equation for fuel, ft^3/ft^3 of waste gas
1	$\frac{1.1\ Cp_5 \Delta T_5 - Cp_2 \Delta T_2 - h_1}{h_3 - 1.1\ Cp_5 \Delta T_5}$
2[a]	$\frac{1.1\ Cp_5 \Delta t_5 - Cp_2 \Delta T_2 - h_1}{h_3 - 1.1\ Cp_5 \Delta T_5}$
3[b,c]	$\frac{1.1\ Cp_5 \Delta T_5 - Cp_2 \Delta T_2 - h_1}{(h_3 - 1.1\ Cp_5 \Delta T_5) - 10.54\ Cp_5\ \Delta T_5\ (1+E)/(1+K)}$
4	$\frac{1.1\ Cp_5 \Delta T_5 + 1.1\ [\Sigma m_i\ (X + Y/4 - Z/2)_i]\ 4.79\ (1+E)\ Cp_5 \Delta T_5 - Cp_2 \Delta T_2 - h_1}{h_3}$
5	Waste gas is premixed and used as fuel.
6	$\frac{Q_3}{Q_2} = \frac{1.1\ [1 + (\Sigma m_i\ (X + Y/4 - Z/2)_i - m_{O_2})\ 4.79]\ Cp_5 \Delta T_5 - [Cp_2 \Delta T_2 + h_1]}{(h_3 - 1.1\ Cp_5 \Delta T_5) - 10.54\ (1+E)\ Cp_5 \Delta T_5/(1+K)}$

[a] The waste gas for this category must be adjusted for dilution (See Table 3B-2).

[b] E - Excess air used

[c] K - Constant = $9.58E/[(h_3/Cp_5 \Delta T_5) - 1]$.

Section 4

CARBON ADSORBERS

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4.1 Process Description

4.1.1 Introduction

In air pollution control, adsorption is employed to remove volatile organic compounds (VOC's) from low to medium concentration gas streams, when a stringent outlet concentration must be met and/or recovery of the VOC is desired. Adsorption itself is a phenomenon where gas molecules passing through a bed of solid particles are selectively held there by attractive forces which are weaker and less specific than those of chemical bonds. During adsorption, a gas molecule migrates to the surface of the solid where it is held by physical atttraction releasing energy--the "heat of adsorption", which approximately equals the heat of condensation. Adsorptive capacity of the solid for the gas tends to increase with the gas phase concentration.

Some gases form actual chemical bonds with the adsorbent surface groups. This phenomenon is termed "chemisorption".

Most gases ("adsorbates") can be removed ("desorbed") from the adsorbent by heating to a sufficiently high temperature, usually via steam, or by reducing the pressure to a sufficiently low value (vacuum desorption). The physically adsorbed species in the smallest pores of the solid and the chemisorbed species may require rather high temperatures to be removed, and for all practical purposes cannot be desorbed during regeneration. For example, approximately 3 to 5 percent of organics adsorbed on virgin activated carbon is either chemisorbed or very strongly physically adsorbed and, for all intents, cannot be desorbed during regeneration.(1)

Adsorbents in large scale use include activated carbon, silica gel, activated alumina, synthetic zeolites, fuller's earth, and other clays. This Section is oriented toward the use of activated carbon, a commonly used adsorbent for VOCs.

4.1.2 Types of Adsorbers

Five types of adsorption equipment are used in collecting gases: (1) fixed regenerable beds; (2) disposable/rechargable cannisters; (3) traveling bed adsorbers; (4) fluid bed adsorbers; and (5) chromatographic baghouses.(2) Of these, the most commonly used in air pollution control are the fixed bed and cannister types.

4.1.2.1 Fixed-bed Units

Fixed-bed units are normally used for controlling continuous, VOC-containing streams over a wide range of flowrates, ranging from several thousand to several hundred thousand cubic feet/minute (cfm). The VOC concentration of streams treated by fixed-bed adsorbers can be as low as several parts per billion (ppbv) by volume in the case of some toxic chemicals or as high as 25% of the VOCs' lower explosive limit (LEL). For most VOCs, this ranges from 2500 to 10,000 ppmv.(3) Fixed-bed adsorbers may be operated in either intermittent or continuous modes. In intermittent operation, the adsorber removes VOC for a specified time (the "adsorption time"), which corresponds to the time during which the controlled source is emitting VOC. After the adsorber and the source are shut down (e.g., overnight), the unit begins the desorption cycle during which the captured VOC is removed from the carbon. This cycle, in turn, consists of three steps: (1) regeneration of the carbon by blowing steam through the bed in the direction opposite to the gas flow; (2) drying of the bed, with compressed air or a fan; and (3) cooling the bed to its operating temperature via a fan. (In some designs, the same fan can be used both for bed drying and cooling.) At the end of the desorption cycle (which usually lasts 1 to 1 1/2 hours), the unit sits idle until the source starts up again.

In continuous operation a regenerated carbon bed is always available for adsorption, so that the controlled source can operate continuously without shut down. For example, two carbon beds can be provided: while one is adsorbing, the second is desorbing/idled. As each bed must be large enough to handle the entire gas flowrate while adsorbing, twice as much carbon must be provided than an intermittent system handling the same flowrate. If the desorption cycle is significantly shorter than the adsorption cycle, it may be more economical to have three, four, or even more beds operating in the system. This can reduce the amount of extra carbon capacity needed or provide some additional benefits, relative to maintaining a low VOC content in the effluent. (See Section 4.2 for a more thorough discussion of this.)

A typical two-bed, continuously operated adsorber system is shown in Figure 4-1. One bed is adsorbing at all times, while the second is desorbing/idled. As shown here, the VOC-laden gas enters vessel #1 through valve A, passes through the carbon bed (shown by the shading) and exits through valve B, from whence it passes to the stack. Meanwhile, vessel #2 is in the desorption cycle. Steam enters through valve C, passes up through the bed and exits through D. The steam-VOC vapor

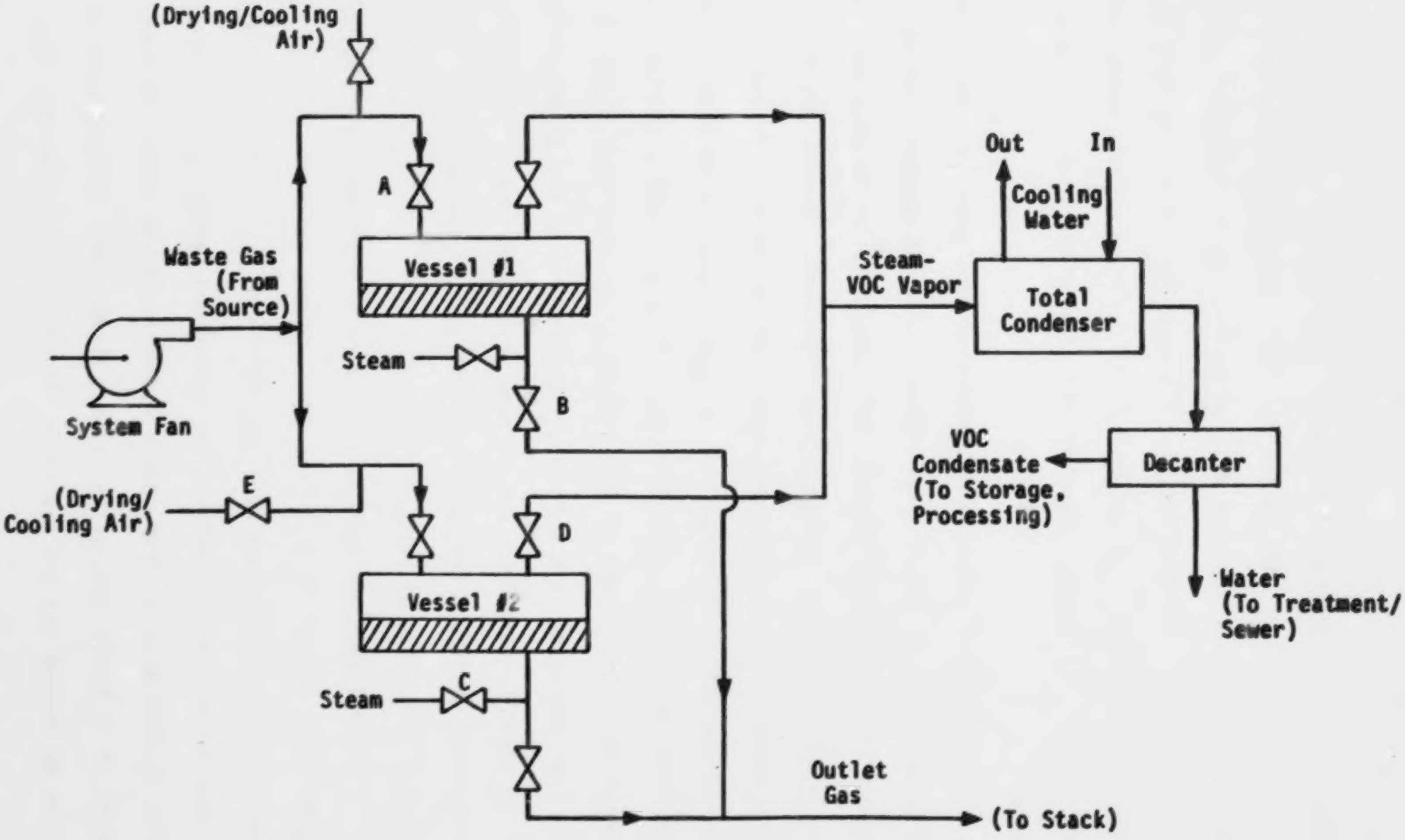
(Drying/Cooling Air)
A
Waste Gas (From Source)
Vessel #1
Steam
B
System Fan
(Drying/ Cooling Air)
E
D
Vessel #2
Steam
C
Steam-VOC Vapor
Out
In
Cooling Water
Total Condenser
Decanter
VOC Condensate (To Storage, Processing)
Water (To Treatment/ Sewer)
Outlet Gas
(To Stack)

Figure 4-1. Typical Two-Bed, Continuously Operated Fixed-Bed Carbon Adsorber System

mixture passes to a condenser, where cooling water condenses the entire mixture. The condensate next passes to a decanter, where the VOC and water layers are separated. The VOC layer is conveyed to storage. If impure, it may receive additional processing, such as distillation. Depending on its quality (i.e., quantity of dissolved organics), the water layer is discharged either to a wastewater treatment facility or to the sewer.

Once steaming is completed, valves C and D are closed and valve E is opened, to allow air to enter to dry and cool the bed. After this is done, the bed is placed on standby until vessel #1 reaches the end of its adsorption cycle. At this time, the VOC-laden gas is valved to vessel #2, while vessel #1 begins its desorption cycle, and the above process is repeated.

In Figure 4-1, the system fan is shown installed ahead of the vessels, though it could also be placed after them. Further, this figure does not show the pumps needed to bring cooling water to the condenser. Nor does it depict the solvent pump which conveys the VOC condensate to storage. Also missing are preconditioning equipment used to cool, dehumidify, or remove particulate from the inlet gases. Such equipment may or may not be needed, depending on the condition of the inlet gas. In any case, preconditioning equipment will not be covered in this Section.

4.1.2.2 Cannister Units

Cannister-type adsorbers differ from fixed-bed units, in that they are normally limited to controlling low-volume, (typically 100 ft^3/min, maximum) intermittent gas streams, such as those emitted by storage tank vents, where process economics dictate that toll regeneration or throw-away cannisters are appropriate. The carbon cannisters may not be desorbed. Alternatively, the carbon may be regenerated at a central facility. Once the carbon reaches a certain VOC content, the unit is shut down, replaced with another, and disposed of or regenerated by the central facility. Each cannister unit consists of a vessel, activated carbon, inlet connection and distributer leading to the carbon bed, and an outlet connection for the purified gas stream.(4) In one design (Calgon's Ventsorb®), 150 lbs of carbon are installed on an 8-inch gravel bed, in a 55-gallon drum. The type of carbon used depends on the nature of the VOC to be treated.

In theory, a cannister unit would remain in service no longer than a regenerable unit would stay in its adsorption cycle. Doing so would prevent the allowable outlet concentration from being exceeded. In reality, however, poor operating practice may result in the cannister remaining connected until the carbon is near saturation. This is because: (1) the carbon (and often the vessel) will probably be disposed of, so there is the temptation to operate it until the carbon is saturated; and (2) unlike fixed-bed units, whose outlet VOC concentrations are monitored continuously (via flame ionization detectors, typically), cannisters are usually not monitored. Thus, the user can only guess at the outlet loading, and could tend to leave a unit in place longer.

4.1.3 Adsorption Theory

At equilibrium, the quantity of gas that is adsorbed on activated carbon is a function of the adsorption temperature and pressure, the chemical species being adsorbed, and the carbon characteristics, such as carbon particle size and pore structure. For a given adsorbent-VOC combination at a given temperature, an adsorption isotherm can be constructed which relates the mass of adsorbate per unit weight of adsorbent ("equilibrium adsorptivity") to the partial pressure of the VOC in the gas stream. The adsorptivity increases with increasing VOC partial pressure and decreases with increasing temperature.

A family of adsorption isotherms having the shape typical of adsorption on activated carbon is plotted in Figure 4-2. This and other isotherms whose shapes are convex upward throughout, are designated "Type I" isotherms. The Freundlich isotherm can be fit to a Type I curve; and it is commonly used in industrial design.(2)

$$w_e = kP^m \qquad (4\text{-}1)$$

where: w_e = equilibrium adsorptivity (lb adsorbate/lb adsorbent)
P = partial pressure of VOC in gas stream (psia)
k,m = empirical parameters

The treatment of adsorption from gas mixtures is complex and beyond the scope of this Section. Except where the VOC in these mixtures have nearly identical adsorption isotherms, one VOC in a mixture will tend displace another on the carbon surface. Generally, VOCs with lower vapor pressures will displace those with higher vapor pressures, resulting in the former displacing the latter previously adsorbed. Thus, during the course of the adsorption cycle the carbon's capacity for a higher vapor pressure constituent decreases. This phenomonen should be considered when sizing the adsorber. To be conservative, one would normally base the adsorption cycle requirements on the least adsorbable component in a mixture and the desorption cycle on the most adsorbable component.(1)

The equilibrium adsorptivity is the maximum amount of adsorbate the carbon can hold at a given temperature and VOC partial pressure. In

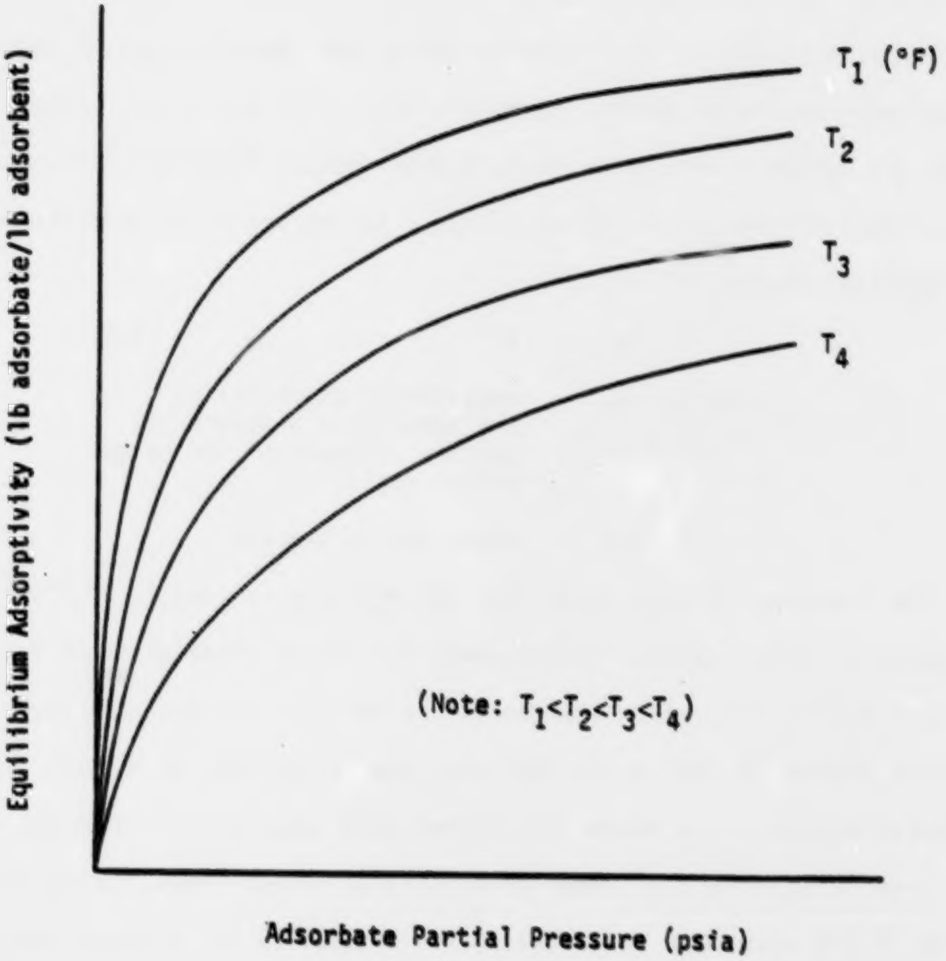

Figure 4-2. Freundlich (Type I) Adsorption Isotherms For Hypothetical Adsorbate

actual control systems, however, the carbon bed is never allowed to reach equilibrium. Instead, once the outlet concentration reaches a preset limit (the "breakthrough concentration"), the adsorber is shut down for desorption or (in the case of cannister units) replacement and disposal. At the point where the vessel is shut down, the bed VOC concentration may only be 50% or less of the equilibrium concentration.

As equation 4-1 indicates, the Freundlich isotherm is a power function that plots as a straight line on log-log paper. Conveniently, for the concentrations/partial pressures normally encountered in carbon adsorber operation, most VOC - activated carbon adsorption conforms to equation 4-1. At very low concentrations, typical of breakthrough concentrations, a linear approximation to the Freundlich isotherm is adequate. However, the Freundlich isotherm does not accurately represent the isotherm at high gas concentrations and thus should be used with care as such concentrations are approached.

Adsorptivity data for several VOCs were obtained from an activated carbon vendor and fitted to the Freundlich equation.(5) These VOCs are listed in Table 4-1. The adsorbates listed include aromatics (e.g., benzene, toluene), chlorinated aliphatics (dichloroethane), and one ketone (acetone). However, the list is far from all-inclusive. Additional isotherm data are available from the activated carbon vendors, handbooks (such as Perry's Chemical Engineer's Handbook), and the literature.

Notice that a range of partial pressures is listed with each set of parameters, "k" and "m". (Note: In one case (m-xylene) the isotherm was so curvilinear that it had to be split into two parts, each with a different set of parameters.) This is the range to which the parameters apply.

Table 4-1 Parameters for Selected Adsorption Isotherms[a,b]

Adsorbate	Adsorption Temp. (°F)	Isotherm Parameters k	m	Range of isotherm[c] (psia)
(1) Benzene	77	0.597	0.176	0.0001-0.05
(2) Chlorobenzene	77	1.05	0.188	0.0001-0.01
(3) Cyclohexane	100	0.508	0.210	0.0001-0.05
(4) Dichloroethane	77	0.976	0.281	0.0001-0.04
(5) Phenol	104	0.855	0.153	0.0001-0.03
(6) Trichloroethane	77	1.06	0.161	0.0001-0.04
(7) Vinyl Chloride	100	2.00	0.477	0.0001-0.05
(8) m-Xylene	77	0.708	0.113	0.0001-0.001
	77	0.527	0.0703	0.001 -0.05
(9) Acrylonitrile	100	0.935	0.424	0.0001-0.015
(10) Acetone	100	0.412	0.389	0.0001-0.05
(11) Toluene	77	0.551	0.110	0.0001-0.05

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[a] Reference 5.

[b] Each isotherm is of the form: $w = kP^m$. (Set text for definition of terms) Data are for adsorption on Calgon type "BPL" carbon (4 x 10 mesh).

[c] Equations should not be extrapolated outside these ranges.

Extrapolation beyond this range--especially at the high end--can introduce inaccuracy to the calculated adsorptivity.

But high-end extrapolation may not be necessary, as the following will show. In most air pollution control applications, the system pressure is approximately one atmosphere (14.696 psia). The upper end of the partial pressure ranges in Table 4-1 goes from 0.04 to 0.05 psia. According to Dalton's Law, at a total system pressure of one atmosphere this corresponds to an adsorbate concentration in the waste gas of 2,720 to 3,400 ppmv. However, as discussed in Section 4.1.2, the adsorbate concentration is usually kept at 25% of the lower explosive limit (LEL). For many VOCs, the LEL ranges from 1 to 1.5 volume %, so that 25% of the LEL would be 0.25 to 0.375% or 2,500 to 3,750 ppmv, which approximates the high end of the partial pressure ranges in Table 4-1.

Finally, each set of parameters applies to a fixed adsorption temperature, ranging from 77° to 104° F. These temperatures reflect typical operating conditions, although adsorption can take place as low as 32°F and even higher than 104°F. As the adsorption temperature increases to much higher levels, however, the equilibrium adsorptivity decreases to such an extent that VOC recovery by carbon adsorption may become economically impractical.

4.2 Design Procedure

4.2.1 Sizing Parameters

Data received from adsorber vendors indicate that the size and purchase cost of a fixed-bed or cannister carbon adsorber system primarily depend on four parameters:

(1) The volumetric flowrate of the VOC laden gas passing through the carbon bed(s);

(2) The inlet and outlet VOC mass loadings of the gas stream;

(3) The adsorption time (i.e., the time a carbon bed remains on-line to adsorb VOC before being taken off-line for desorption of the bed);

(4) The working capacity of the activated carbon.

In addition, the cost could also be affected by other stream conditions, such as the presence/absence of excessive amounts of particulate, moisture, or other substances which would require the use of extensive pretreatment and/or corrosive-resistant construction materials.

The purchased cost depends to a large extent on the volumetric flowrate (usually measured in actual ft^3/min). The flowrate, in turn, determines the size of the vessels housing the carbon, the capacities of the fan and motor needed to convey the waste gas through the system, and the diameter of the internal ducting.

Also important are the VOC inlet and outlet gas stream loadings, the adsorption time, and the working capacity of the carbon. These variables determine the amount and cost of carbon charged to the system initially and, in turn, the cost of replacing that carbon after it is exhausted (typically, five years after startup). Moreover, the amount of the carbon charge affects the size and cost of the auxiliary equipment (condenser, decanter, bed drying/cooling fan), because the sizes of these items are tied to the amount of VOC removed by the bed. The amount of carbon also has a bearing on the size and cost of the vessels.

To illustrate this effect, for each of a range of flowrates, the VOC inlet concentration was increased ten-fold from 500 to 5000 ppm, while the respective flowrates and adsorption times were held constant. The resulting purchased costs obtained from a vendor increased by an average of 27%.(6) Part of these increases was needed to pay for the additional carbon required. However, some was also needed for enlarging the adsorber vessels to accomodate the added carbon and for the additional structural steel needed to support the larger vessels. Also, larger condensers, decanters, cooling water pumps, etc., were necessary to treat the more concentrated VOC streams. (See Section 4.3.)

The VOC inlet loading is set by the source parameters, while the outlet loading is set by the VOC emission limit. (For example, in many states, the average VOC outlet concentration from adsorbers may not exceed 25 ppm.)

4.2.2 Determining Adsorption and Desorption Times

The relative times for adsorption and desorption and the adsorber bed configuration (i.e., whether single or multiple and series or parallel adsorption beds are used) establish the adsorption/desorption cycle profile. The cycle profile is important in determining carbon and vessel requirements and in establishing desorption auxiliary equipment and utility requirements. An example will illustrate. In the simplest case, an adsorber would be controlling a process which emits a relatively small amount of VOC intermittently--say, during one 8-hour shift per day. During the remaining 16 hours the system would either be desorbing or on stand-by. Such a system would only require a single bed, which would contain enough carbon to treat eight hours worth of gas flow at the specified inlet concentration, temperature, and pressure. Multiple beds, operating in parallel, would be needed to treat large gas flows (>100,000 std. ft^3/min, generally)(6), as there are practical limits to the sizes to which adsorber vessels can be built. But, regardless of whether a single bed or multiple beds were used, the system would only be on-line for part of the day.

However, if the process were operating continuously, extra carbon capacity would have to be built into the system. The amount of this extra capacity would depend on the number of carbon beds that would be adsorbing at any one time, the length of the adsorption period relative to the desorption period, and whether the beds were operating in parallel or in series. If one bed were adsorbing, a second would be needed to come on-line when the first was shut down for desorption. In this case, 100% extra capacity would be needed. Similarly, if five beds in parallel were adsorbing at any given time, again only one extra bed would be needed and the extra capacity would be 20% (i.e., 1/5)--provided, of course, that the adsorption time were at least five times as long as the desorption time. The relationship between adsorption time, regeneration time, and the required extra capacity can be generalized.

$$M_C = M_{CI} \times f \qquad (4\text{-}2)$$

where: M_C, M_{CI} = amounts of carbon required for continuous or intermittent control of a given source, respectively (lbs)

f = extra capacity factor (dimensionless)

The factor, f, is related to the number of beds adsorbing (N_A) and desorbing (N_D) in a continuous system as follows:

$$f = 1 + {}^{N_D}/N_A \qquad (4\text{-}3)$$

(Note: N_A is also the number of beds in an intermittent system that would be adsorbing at any given time. The total number of beds in the system would be $N_A + N_D$.)

It can be shown that the number of desorbing beds required in a continuous system (N_D) is related to the desorption time (O_D), adsorption time (O_A), and the number of adsorbing beds, as follows:

$$\theta_D \le \theta_A (N_D / N_A) \qquad (4-4)$$

(Note: θ_D is the total time needed for bed regeneration, drying, and cooling.) For instance, for an eight-hour adsorption time, in a continuously operated system of seven beds (six adsorbing, one desorbing) θ_D would have to be 1 1/3 hrs or less (8 hrs/6 beds). Otherwise, additional beds would have to be added to provide sufficient extra capacity during desorption.

4.2.3 Estimating Carbon Requirement

4.2.3.1 Carbon Estimation Procedures Developed

Obtaining the carbon requirement (M_c or M_{c_I}) is not as straightforward as determining the other adsorber design parameters. When estimating the carbon charge, the depth of the approach used depends on the data and calculational tools available. In preparing this Section of the Manual, we have developed two procedures for estimating the carbon requirement. The first procedure, described in more detail below, is based to a large extent on rules-of-thumb. This procedure, sometimes employed by adsorber vendors, is relatively simple and easy to use, though it normally yields results incorporating a large safety margin.

The second procedure yields a more accurate estimate of the carbon requirement but requires additional input data. In addition, the procedure is of such mathematical complexity that a microcomputer is needed to use it. This procedure is centered on the "BED_SIZE" model, developed by William L Klotz of Research Triangle Institute. The program "...uses a detailed mathematical description of the adsorption process to predict the bed size needed to maintain the effluent [outlet] concentration below a defined allowable maximum when the inlet concentration is constant over the time period for operation."(7) Inputs needed to run the program include the inlet and desired outlet ("breakthrough") VOC concentrations, gas stream temperature, and the "breakthrough time," which is usually equal to the adsorption time. Other inputs needed are the carbon particle size and bed void fraction and the gas superficial velocity through the bed. (Superficial velocity is discussed in Section 4.3.1.) As outputs, the program provides the carbon equilibrium capacity, working capacity (see below), the required bed size (as measured by the bed depth) and pressure drop, and the bed efficiency--"the percent removal of all adsorbate entering the bed up to the breakthrough time." This efficiency is calculated by integrating the ratio of VOC removed to the VOC inlet rate over the entire adsorption cycle. (See Section 4.4.3. for more on this.)

These BED_SIZE outputs, along with the bed superficial velocity and other parameters, are input to a second model ("CARADS"), developed by the Economic Analysis Branch. This model is used to determine the carbon requirement, size the adsorber vessels, and estimate the system capital and annual costs in a manner similar to the rule-of-thumb procedure detailed in the next sections. However, where the rule-of-thumb procedure may be done by hand, a microcomputer is needed to utilize the CARADS model.

4.2.3.2. Carbon Estimation Procedure Used in Manual

The rule-of-thumb carbon esimation procedure is based on the "working capacity" (w_c, lb VOC/lb carbon). This is the difference per unit mass of carbon between the amount of VOC on the carbon at the end of the adsorption cycle and the amount remaining on the carbon at the end of the desorption cycle. It should not be confused with the "equilibrium capacity" (w_e) defined above in Section 4.1.3. Recall that the equilibrium capacity measures the capacity of virgin activated carbon when the VOC has been in contact with it (at a constant temperature and partial pressure) long enough to reach equilibrium. In adsorber design, it would not be feasible to allow the bed to reach equilibrium. If it were, the outlet concentration would rapidly increase beyond the allowable outlet (or "breakthrough") concentration until the outlet concentration reached the inlet concentration. During this period the adsorber would be violating the emission limit.

The working capacity is some fraction of the equilibrium capacity. Table 4-2 compiles working capacities for selected VOCs. These data were obtained from an adsorber manufacturer.(8) For comparison, Table 4-2 also shows the equilibrium capacities for some of these VOCs, which were computed from the parameters given above in Table 4-1 in Section 4.1. Note that the working capacities range from 24 to 100% of the respective capacities at equilibrium, with an average value of 48%. This average approximates the rule-of-thumb used by adsorber vendors--that is, working capacity equals 50% of equilibrium capacity. Further, like the equilibrium adsorptivity, the working capacity depends upon the temperature, the VOC partial pressure, and the VOC composition. The working capacity also depends on the flow rate and the carbon bed parameters.

The working capacity, along with the adsorption time and VOC inlet loading, is used to compute the carbon requirement for a cannister adsorber or for an intermittently operated fixed-bed adsorber as follows:

Table 4-2. Equilibrium vs. Working Carbon Capacity for Selected VOCs(8)

Compound	Adsorption Temp. (°F)[a]	Partial Press. (psia)[b]	Carbon Capacities[c]		w_c/w_e Ratio
			Equilibrium (w_e)	Working (w_c)	
(1) Acetone	100	0.0147	0.0798	0.08	1.0
(2) Benzene	77	0.000147	0.126	0.06	0.48
(3) Cyclohexane	100	0.00441	0.163	0.06	0.37
(4) Toluene	77	0.00294	0.290	0.07	0.24
(5) m-Xylene[d]	77	0.00147	0.333	0.10	0.30
				Average	0.48

[a] Temperature at which equilibrium capacity was measured.

[b] Directly proportional to the VOC concentration in the influent (ppmv). Calculated as follows:
Partial pressure = Conc. (ppmv) x 14.696 psia x 10^{-6}

[c] Measured as lb of VOC/lb. of carbon. Equilibrium capacities were calculated using parameters in Table 4-1; working capacities were obtained from Reference 8.

[d] Actually, the compound listed in Reference 8 was "xylene." The precise structure -- i.e., ortho, meta, or para -- was not given. However, as the molecular weight of all three is the same, we have assumed that the working capacity data would apply to m-xylene, as well as to the other two isomers.

$$M_{cI} = \frac{m_{VOC}\ \theta_A}{w_c} \qquad (4\text{-}5)$$

where: m_{VOC} = VOC inlet loading (lb/hr)

Combining this with equations (4-2) and (4-3) yields the general equation for estimating the system total carbon charge:

$$M_c = \frac{m_{VOC}\ \theta_A}{w_c}\ (1 + N_D/N_A) \qquad (4\text{-}6)$$

Values for w_c may be obtained from Table 4-2 or similar data sources. If no value for w_c is available for the VOC (or VOC mixture) in question, the working capacity may be estimated at 50% of the equilibrium capacity, as follows:

$$w_c \cong 0.5\ w_e(\text{max}) \qquad (4\text{-}7)$$

where: w_e(max) = the equilibrium capacity (lb VOC/lb carbon) taken at the adsorber inlet (i.e., the point of maximum VOC concentration).

(Note: To be conservative, this 50% figure should be lowered if short desorption cycles, very high vapor pressure consituents, or difficult-to-desorb VOCs are involved.)

* * *

Example: A source emitting 100 lb/hr toluene is to be collected in a carbon adsorber. The system operates continuously, with two beds adsorbing at all times. For convenience, adsorption and desorption times of 12 and 1.5 hours, respectively, have been chosen. The total gas flowrate is 35,000 ft^3/min at 77°F and 1 atmosphere (or approximately 200 ppmv of VOC). Assume negligible quantities of particulate and moisture. Calculate the system carbon requirement. Solution: From Table 4-2, we find the working capacity of toluene to be 0.07 lb/lb carbon. Since the regeneration time

(1.5 hrs) is less than 1/2 the adsorption time (see eq. 4-4), one extra bed will be adequate. Substituting these values into equation 4-6 yields the carbon requirement:

$$M_C = \frac{(100 \text{ lb VOC/hr})(12 \text{ hr})}{0.07 \text{ lb VOC/lb carbon}} (1 + 1/2)$$

$$= \text{approximately } 25{,}700 \text{ lbs.}$$

* * *

As equation 4-6 shows, the carbon requirement is directly proportional to the adsorption time. This would tend to indicate that a system could be designed with a shorter adsorption time to minimize the carbon requirement (and purchased cost). There is a trade-off here not readily apparent from equation 4-6, however. Certainly, a shorter adsorption time would require less carbon. But, it would also mean that a carbon bed would have to be desorbed more frequently. This would mean that the regeneration steam would have to be supplied to the bed(s) more frequently to remove (in the long run) the same amount of VOC. Further, each time the bed is regenerated the steam supplied must heat the vessel and carbon, as well as drive off the adsorbed VOC. And the bed must be dried and cooled after each desorption, regardless of the amount of VOC removed. Thus, if the bed is regenerated too frequently, the bed drying/cooling fan must operate more often, increasing its power consumption. Also, more frequent regeneration tends to shorten the carbon life. As a rule-of-thumb, the optimum regeneration frequency for fixed-bed adsorbers treating streams with moderate to high VOC inlet loadings is once every 8 to 12 hours.(1)

4.3 Estimating Total Capital Investment

An entirely different procedure should be used in estimating the purchased costs of fixed-bed and cannister-type adsorbers. Therefore, they will be discussed separately.

4.3.1 Fixed-Bed Systems

As indicated in the previous section, the purchased cost is a function of the volumetric flowrate, VOC inlet and outlet loadings, the adsorption time, and the working capacity of the activated carbon. As Figure 4-1 shows, the adsorber system is made up of several different items. Of these, the adsorber vessels and the carbon comprise nearly 3/4 of the purchased cost. There is also auxiliary equipment, such as fans, pumps, condensers, decanters, and internal piping. But because these usually comprise a small part of the total purchased cost, they may be "factored" from the costs of the two major items without introducing significant error. The costs of these major items will be considered separately.

4.3.1.1 Carbon Cost

This cost (C_c,\$) is simply the product of the initial carbon requirement (M_c) and the current price of carbon. As adsorber vendors buy carbon in very large quantities (million-pound lots or larger), their cost is somewhat lower than the list price. Current vendor costs typically range from \$1.60 to \$2.00/lb (spring 1986 dollars). Taking the midpoint of this range, we have:

$$C_c = 1.80\ M_c \qquad (4\text{-}8)$$

4.3.1.2 Vessel Cost

The cost of an adsorber vessel is primarily determined by its dimensions which, in turn, depend upon the amount of carbon it must hold and the superficial gas velocity through the bed that must be maintained for optimum adsorption. The desired superficial velocity is used to calculate the cross-sectional area of the bed perpendicular to the gas flow. An acceptable superficial velocity is established empirically, considering desired removal efficiency, the carbon particle size and bed porosity, and other factors. For example, one adsorber vendor recommends a superficial bed velocity of 85 ft/min(6), while an activated carbon manufacturer cautions against exceeding 60 ft/min in systems operating at one atmosphere.(5) Another vendor uses a 65 ft/min. superficial face velocity in sizing its adsorber vessels.(9) Lastly, there are practical limits to vessel dimensions which also influence their sizing. That is, due to shipping restrictions, vessel diameters rarely exceed 12 feet, while their length is generally limited to 50 feet.(9)

The cost of a vessel is usually correlated with its weight. However, as the weight is often difficult to obtain or calculate, the cost may be determined from the external surface area. This is true because the vessel material cost -- and the cost of fabricating that material -- is directly proportional to its surface area. The surface area (S, ft^2) of a vessel is a function of its length (L) and diameter (D), which in turn, depend upon the superficial bed face velocity, the L/D ratio, and other factors.

Most commonly, adsorber vessels are cylindrical in shape and erected horizontally (as in Figure 4-1). In these cases, it can be shown that:

$$D = \frac{0.127 M_c' v_b}{Q} \qquad (4\text{-}9)$$

and:

$$L = \frac{7.87}{M_c}(Q'/v_b)^2 \qquad (4\text{-}10)$$

where: v_b = bed superficial velocity (ft/min)

M_c' = carbon requirement per vessel (lbs)

Q' = volumetric flowrate per adsorbing vessel (ft^3/min)

Equations (4-9) and (4-10) also assume that the carbon occupies 1/3 of the vessel volume(6,9) and that the carbon's bulk density is 30 lb/ft^3.

Finally, for a cylinder:

$$S = \pi D(L + D/2) \qquad (4\text{-}11)$$

Similar equations can be developed for other vessel shapes, configurations, etc.

Based on vendor data, we developed a correlation between adsorber vessel cost and surface area:(9)

$$C_v = \exp\left[18.827 - 3.3945 \ln S + 0.3090 (\ln S)^2\right] \qquad (4\text{-}12)$$

where: C_v = vessel cost (spring 1986 \$), F.O.B. vendor

and: $228 \leq S \leq 2{,}111\ ft^2$

These units would be made of 304 stainless steel, which is the most common material used in fabricating adsorber vessels.(6,9) However, to obtain the cost of a vessel fabricated of another material, multiply C_v by an adjustment factor (F_m). A few of these factors are listed below:

Material	F_m Factor	Reference(s)
Stainless steel, 316	1.3	6, 9, 10
Carpenter 20 CB-3	1.9	10
Monel-400	2.3	6, 10
Nickel-200	3.2	10
Titanium	4.5	10

4.3.1.3. Total Purchased Cost

As stated earlier, the costs of such items as the fans, pumps, condenser, decanter, instrumentation, and internal piping can be factored from the sum of the costs for the carbon and vessels. Based on four data points derived from costs supplied by an equipment vendor (9), we found that, depending on the total gas flowrate (Q), the cost of the carbon and vessels together comprised from 50% to 85% of the total adsorber equipment cost. These data points spanned a gas flowrate range of 4,000 to 500,000 scfm. The average of these points was 72%, with a standard deviation of 16%. Taking the reciprocal of this average (i.e., 1/0.72 = 1.39), we can write:

$$C_A = 1.39[C_c + C_v] \tag{4-13}$$

where: C_c = cost of carbon

C_v = cost of vessel(s)

C_A = cost of adsorber equipment

4.3.1.4. Total Capital Investment

As discussed in Section 2, in the methodology used in this Manual, the total capital investment (TCI) is estimated from the total purchased cost via an overall direct/indirect installation cost factor. A breakdown of that factor for carbon adsorbers is shown in Table 4-3. As Section 2 indicates, the TCI also includes costs for land, working capital, and off-site facilities, which are not included in the direct/indirect installation factor. However, as these items are rarely required with adsorber systems, they will not be considered here. Further, no factors have been provided for site preparation (S.P.) and buildings (Bldg.), as these site-specific costs depend very little on the purchased equipment cost.

Note that the installation factor is applied to the total purchased equipment cost, which includes the cost of the stack and external ductwork and such costs as freight and sales taxes (if applicable). ("External ductwork" is that ducting needed to convey the exhaust gas from the source to the adsorber system, and then from the adsorber to the stack. Costs for ductwork

Table 4-3. Direct and Indirect Installation Factors for Carbon Adsorbers[a]

Cost Item	Cost Factor (Fraction of Indicated Cost)
DIRECT COSTS	
1) Purchased equipment cost	
Adsorber	As required)
Auxiliary equipment[b]	As required) A
Instruments and controls[c]	0.1 A
Taxes	0.03 A
Freight	0.05 A
Total Purchased Equipment cost	B(=1.18A)
2) Installation direct costs	
Foundations and supports	0.08 B
Erection and handling	0.14 B
Electrical	0.04 B
Piping	0.02 B
Insulation	0.01 B
Painting	0.01 B
Site preparation (S.P.), Buildings (Bldg.)	As required
Total Installation Direct Costs	0.30 B + S.P. + Bldg.
Total Direct Cost	1.30B + S.P. + Bldg.
INDIRECT COSTS	
Engineering and supervision	0.10 B
Construction and field expenses	0.05 B
Construction fee	0.10 B
Start-up	0.02 B
Performance test	0.01 B
Contingency	0.03 B
Total Indirect Cost	0.31B
Total Direct and Indirect Costs = Total Capital Investment	1.61B + S.P. + Bldg.

[a] Reference 11.

[b] Includes external ductwork, stack, and any other equipment normally not included with unit furnished by adsorber vendor.

[c] Instrumentation cost usually included in cost of vendor-supplied adsorber unit.

and stacks are shown elsewhere in this Manual.) Normally, the adjustment would also cover the instrumentation cost, but this cost is usually included with the adsorber purchased cost. Finally, note that these factors reflect "average" installation conditions and could vary considerably, depending upon the installation circumstances.

4.3.2 Cannister Systems

Once the carbon requirement is estimated using the above procedure, the number of cannisters is determined. This is done simply by dividing the total carbon requirement (M_c) by the amount of carbon contained by each cannister (typically, 150 lbs.). This quotient, rounded to the next highest digit, yields the required number of cannisters to control the vent in question.

Costs for a typical cannister (Calgon's Ventsorb®) are listed in Table 4-4. These costs include the vessel, carbon, and connections, but do not include freight or installation charges. Note that the cost per unit decreases as the quantity purchased increases. Each cannister contains Calgon's "BPL" carbon (4 x 10 mesh), which is commonly used in industrial adsorption. However, to treat certain VOCs, more expensive speciality carbons (e.g., "FCA 4 x 10") are needed. These carbons can increase the equipment cost by 60% or more.(4)

As fewer installation materials and labor are required to install a cannister unit than a fixed-bed system, the composite installation factor is consequently lower. The only costs required are those needed to place the cannisters at, and connect them to, the source. This involves a small amount of piping only; little or no electrical work, painting, foundations, or the like would be needed. Twenty percent of the sum of

Table 4-4 Equipment Costs (Spring 1986 $) for a Typical Cannister Adsorber[a]

Quantity	Equipment Cost (each)[b]
1-3	$687
4-9	659
10-29	622
≥30	579

[a] Reference 4.

[b] These costs are F.O.B., Pittsburgh, PA. They do not include taxes and freight charges.

the cannister(s) cost, freight charges, and applicable sales taxes would cover this installation cost.

4.4. Estimating Total Annual Cost

As Section 2 of this *Manual* explains, the annual operating cost is comprised of three components: *direct costs*, *indirect costs*, and *recovery credits*. These will be considered separately.

4.4.1 Direct Annual Costs

These include the following expenditures: steam, cooling water, electricity, carbon replacement, operating and supervisory labor, and maintenance labor and materials. Of these, only *electricity* and *solid waste disposal* would apply to the cannister-type adsorbers.

4.4.1.1 Steam:

As explained in Section 4.1, steam is used during the desorption cycle. The quantity of steam required will depend on the amount of carbon in the vessel, the vessel dimensions, the type and amount of VOC adsorbed, and other variables. Experience has shown that the steam requirement ranges from approximately *3 to 4* lbs of steam/lb of adsorbed VOC.(6,9) Using the midpoint of this range, we can develop the following expression for the annual steam cost:

$$C_s = 3.50\text{x}10^{-3}\ m_{VOC}\ \theta s\ ps \qquad (4\text{-}14)$$

where: C_s = steam cost (\$/yr)

θ_s = system operating hours (hr/yr)

m_{VOC} = VOC inlet loading (lbs/hr)

ps = steam price (\$/thous. lbs)

If steam price data are unavailable, one can estimate its cost at *120%* of the fuel cost. For example, if the local price of natural gas were \$5.00/million BTU, the estimated steam price would be \$6.00/million BTU = \$6.00/thousand lbs. (The 20% factor covers the capital and annual costs of producing the steam.)

4.4.1.2 Cooling Water:

Cooling water is consumed by the condenser in which the steam-VOC mixture leaving the desorbed carbon bed is totally condensed. Most of the condenser duty is comprised of the latent heat of vaporization (ΔH_v) of the steam and VOC. As the VOC ΔH_v are usually small compared to the steam ΔH_v (about 1000 BTU/lb), the VOC ΔH_v may be ignored. So may the sensible heat of cooling the water-VOC condensate from the condenser inlet temperature (about 212°F) to the outlet temperature. Therefore, the cooling water requirement is essentially a function of the steam usage and the allowable temperature rise in the coolant, which is typically 30° to 40°F(6). Using the average temperature rise (35°F), we can write:

$$C_{cw} = 3.43 C_s \, p_{cw} \qquad (4\text{-}15)$$

where: C_{cw} = cooling water cost (\$/yr)

p_{cw} = cooling water price (\$/thous. gal.)

If the cooling water price is unavailable, use \$0.15 to \$0.30/thousand gallons.

4.4.1.3 Electricity:

In fixed-bed adsorbers, electricity is consumed by the system fan, bed drying/cooling fan, cooling water pump, and solvent pump(s). Both the system and bed fans must be sized to overcome the pressure drop through the carbon beds. But, while the system fan must continuously convey the total gas flow through the system, the bed cooling fan is only used during a part of the desorption cycle (one-half hour or less).

For both fans, the horsepower needed depends both on the gas flowrate and the pressure drop through the carbon bed. The pressure drop through the bed (ΔP_b) depends on several variables, such as the adsorption temperature, bed velocity, bed characteristics (e.g., void fraction), and thickness. But, for a given temperature and carbon, the pressure drop per unit thickness depends solely on the gas velocity. For instance, for Calgon's "PCB" carbon (4 x 10 mesh), the following relationship holds:(5)

$$\Delta P_b/t_b = 0.03679\ v_b + 1.107 \times 10^{-4}\ v_b^2 \qquad (4\text{-}16)$$

where: $\Delta P_b/t_b$ = pressure drop through bed (inches of water/foot of carbon)

v_b = superficial bed velocity (ft/min)

The bed thickness (t_b,ft) is the quotient of the bed volume (V_b) and the bed cross-sectional area (A_b). For a 30 lb/ft^3 carbon bed density, this becomes:

$$t_b = \frac{V_b}{A_b} = \frac{0.0333M_c'}{A_b} \qquad (4\text{-}17)$$

(For instance, for horizontally-erected cylindrical vessels, $A_b \cong LD$.)

Once ΔP_b is known, the system fan horsepower requirement (hp_{sf}) can be calculated:

$$hp_{sf} = 2.50 \times 10^{-4}\ Q\ \Delta P_s \qquad (4\text{-}18)$$

where: Q = gas volumetric flowrate through system (ft^3/min)

ΔP_s = total system pressure drop = $\Delta P_b + 1$

(The extra inch accounts for miscellaneous pressure losses through the external ductwork and other parts of the system.(6))

This equation incorporates a fan efficiency of 70% and a motor efficiency of 90%, or 63% overall.

The horsepower requirement for the bed drying/cooling fan (hp_{cf}) is computed similarly. While the bed fan pressure drop would still be ΔP_b, the gas flowrate and operating times would be different. (For example, to cool a bed from the regeneration temperature to the adsorption temperature, would require from 3 to 3.5 SCFM/lb of carbon, based on a 30-minute cooling/drying time.)

The cooling water pump horsepower requirement (hp_{cwp}) would be computed as follows:

$$hp_{cwp} = \frac{2.52 \times 10^{-4}\, q_{cw} Hs}{\eta} \qquad (4\text{-}19)$$

where: q_{cw} = cooling water flowrate (gal/min)

H = required head (normally 100 feet of water)

s = specific gravity of fluid relative to water at 60°F.

η = combined pump-motor efficiency

Equation 4-18 may also be used to compute the solvent pump horsepower requirement. In the latter case, the flowrate (q_s) would be different, of course, although the same head--100 ft. of water--could be used. The specific gravity would depend on the composition and temperature of the condensed solvent. For example, the specific gravity of toluene at 100°F would be approximately 0.86 at 70° F. (However, the solvent pump horsepower is usually very small--usually $\leq$0.1 hp.--so its electricity consumption can usually be neglected.)

Once the various horsepowers are calculated, the electricity usage (in kWh) is calculated, by multiplying each horsepower value by 0.746 (the factor for converting hp to kilowatts) and the number of hours each fan or pump operates annually. For the system fan and cooling water pump, the hours would be the annual operating hours for the system (Θs).

But for the bed drying/cooling fan, the time (Θ_b) would be much shorter. This is dependent on the time the bed fan is run during a desorption cycle, the number of desorption cycles per year, the adsorption time (Θ_A), and other factors. For instance, for a three-bed system, with a 12-hour adsorption time and a 0.5-hour bed cooling/drying time per desorption cycle, the bed fan would operate 8.33 hrs. for every 100 hrs of on-stream (system) time.

To obtain the annual electricity cost, simply multiply kWh by the electricity price (in \$/kWh) that applies to the facility being controlled

For cannister units, use equation 4-17 to calculate the fan horsepower requirement. However, instead of ΔP_b, use the following to compute the total cannister pressure drop (ΔPc, inches of water):(4)

$$\Delta P_c = 0.0471Q_c + 9.29 \times 10^{-4}Q_c^2 \qquad (4\text{-}20)$$

where: Q_c = flowrate through the cannister (ft^3/min)

4.4.1.4. Carbon Replacement:

As discussed above, the carbon has a different economic life than the rest of the adsorber system. Therefore, its replacement cost must be calculated separately. Employing the procedure detailed in Section 2, we have:

$$CRC_c = CRF_c\ (1.08C_c + C_{cl}) \qquad (4\text{-}21)$$

where: CRF_c = capital recovery factor for the carbon

1.08 = taxes and freight factor

C_c, C_{cl} = initial cost of carbon and carbon replacement labor cost, respectively (\$/yr) (F.O.B. vendor)

The replacement labor cost covers the labor cost for removing spent carbon from vessels and replacing it with virgin or regenerated carbon. The cost would vary with amount of carbon being replaced, the labor rates, and other factors. For example, to remove and replace a 50,000 - pound charge would require about 16 person-days, which, at typical wage rates, is equivalent to approximately $0.05/lb replaced.(12)

A typical life for the carbon is five years. However, if the inlet contains VOCs that are very difficult to desorb, tend to polymerize, or react with other constituents, a shorter carbon lifetime--perhaps as low as two years--would be likely.(1) For a five-year life and 10% interest rate, $CRF_c = 0.2638$.

4.4.1.5 Solid Waste disposal:

Disposal costs are rarely incurred with fixed-bed adsorbers, because the carbon is almost always regenerated in place, not discarded. In certain cases, the carbon in cannister units is also regenerated, either off-site or at a central regeneration facility on-site. However, most cannister adsorbers are disposed of once they become saturated. The entire cannister--carbon, drum, connections, etc.--is shipped to a secure landfill. The cost of landfill disposal could vary considerably, depending on the number of cannisters disposed of, the location of the landfill, etc. Based on data obtained from two large landfills, for instance, the disposal cost would range from approximately $35 to $65 per cannister excluding transportation costs.(13,14)

4.4.1.6 Operating and Supervisory Labor:

The operating labor for adsorbers is relatively low, as most systems are automated and require little attention. One-half operator hour per shift is typical.(11) Add to this 15% to cover supervisory labor, as Section 2 suggests. The annual labor cost would then be the product of these labor requirements and their respective wage rates ($/hr) which, naturally, would vary according to the facility location, type of industry, etc.

4.4.1.7 Maintenance Labor and Materials:

Use 0.5 hours/shift for maintenance labor (11) and the applicable maintenance wage rate. If the latter data are unavailable, estimate the maintenance wage rate at 110% of the operating labor rate, as Section 2 suggests. Finally, for maintenance materials, add an amount equal to the maintenance labor, also per Section 2.

4.4.2 Indirect Annual Costs

These include such costs as capital recovery, property taxes, insurance, overhead, and administrative costs ("G&A"). The capital recovery cost is based on the equipment lifetime and the annual interest rate employed. (See Section 2 for a thorough discussion of the capital recovery cost and the variables that determine it.) For adsorbers, the system lifetime is typically ten years, except for the carbon, which, as stated above, typically needs to be replaced after five years. Therefore, when figuring the system capital recovery cost, one should base it on the installed capital cost less the cost of replacing the carbon (i.e., the carbon cost plus the cost of labor necessary to replact it). Substituting the initial carbon and replacement labor costs from equation 4-21, we obtain:

$$CRC_s = [TCI - (1.08C_c + C_{cl})]\ CRFs \qquad (4\text{-}22)$$

where: CRC_s = capital recovery cost for adsorber system ($/yr)

TCI = total capital investment ($)

1.08 = taxes and freight factor

C_{cl} = initial carbon cost (F.O.B. vendor) and carbon replacement cost, respectively ($)

CRF_s = capital recovery factor for adsorber system (defined in Section 2).

For a ten-year life and a 10% annual interest rate, the CRF_s would be 0.1628.

As Section 2 suggests, the suggested factor to use for property taxes, insurance, and administrative charges is 4% of the TCI. Finally, the overhead is calculated as 60% of the sum of operating, supervisory, and maintence labor, and maintenance materials.

* * *

The above procedure applies to cannister units as well, except that, in most cases, the carbon is not replaced--the entire unit is. Cannisters are generally used in specialized applications. The piping and ducting cost can usually be considered a capital investment with a useful life of ten years. However, whether the cannister itself would be treated as a capital or an operating expense would depend on the particular application and needs to be evaluated on a case-by-case basis.

4.4.3 Recovery Credits

These could apply to the VOC which is adsorbed, then desorbed, condensed, and separated from the steam condensate. However, if the VOC layer contained impurities or were a mixture of compounds, it would require further treatment, such as distillation. Purification and separation costs are beyond the scope of this Section. Suffice it to say that the costs of these operations could offset and, if large enough, obliterate any recovery credits. In any case, the following equation can be used to calculate these credits:

where: P_{VOC} = resale value of the captured VOC ($/lb)

RC = recovery credit ($/yr)

E = adsorber VOC control efficiency

By definition, the efficiency (E) is the difference between the inlet and outlet VOC mass loadings, divided by the inlet loading. However, during an adsorption cycle the outlet VOC loading will increase from essentially zero at the start of the cycle to the breakthrough concentration at the end of the cycle. Because the efficiency is a function of time, it should be calculated via integration over the length of the adsorption cycle. To do this would require knowledge of the temporal variation of the outlet loading during the adsorption cycle. If this knowledge is not available to the Manual user, a conservative approximation of the efficiency may be made by setting the outlet loading equal to the breakthrough concentration.

4.4.4 Total Annual Cost

Finally, as explained in Section 2, the total annual cost (TAC) is the sum of the direct and indirect annual costs, less any recovery credits, or:

$$TAC = DC + IC - RC \qquad (4\text{-}24)$$

References for Section 4

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Section 5

FABRIC FILTERS

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5.1 Process Description

5.1.1 Introduction

A fabric filter unit consists of one or more isolated compartments containing rows of fabric filter bags or tubes. Particle-laden gas passes up (usually) along the surface of the bags then radially through the fabric. Particles are retained on the upstream face of the bags while the cleaned gas stream is vented to the atmosphere. The filter is operated cyclically alternating between relatively long periods of filtering and short periods of cleaning. During cleaning, dust that has accumulated on the bags is removed from the fabric surface and deposited in a hopper for subsequent disposal.

This device will collect particle sizes ranging from submicron to several hundred microns in diameter at efficiencies generally in excess of 99 or 99.9 percent. The dust cake collected on the fabric is primarily responsible for such high efficiency. Gas temperatures up to 500 °F, with surges to 550 °F can be accommodated routinely. Most of the energy used to operate the system appears as pressure drop across the bags and associated hardware and ducting. Typical values range from 5 to 20 in. of water gauge. Fabric filters are used where high-efficiency particle collection is required. Limitations are imposed by gas characteristics (temperature and corrosivity) and particle characteristics (primarily stickiness) that affect the fabric or its operation and that cannot be accommodated economically.

Important process variables include particle characteristics, gas characteristics, and fabric properties. The most important design parameter is the air- or gas-to-cloth ratio (volumetric flow rate divided by fabric area, $(ft^3/min)/ft^2$), and the usual operating parameter of interest is pressure drop across the filter system. The major distinguishing operating feature of fabric filters is the ability to renew the filtering surface periodically by cleaning.

5.1.2 Types of Fabric Filters

Fabric filters can be categorized by several means, including type of cleaning (shaker, reverse-air, pulse-jet), direction of gas flow (from inside the bag towards the outside or vice versa), location of the system fan (suction or pressure), or size (low, medium, or high gas flow quantity). Cleaning methods are discussed more fully in this section, and the other categories are described in Section 5.2.

5.1.2.1 Shaker Cleaning

For any type of cleaning, enough energy must be imparted to the fabric to overcome the adhesion forces holding dust to the bag. In shaker cleaning, used with inside to outside gas flow, this is accomplished by suspending the bag from a motor-driven hook or framework that oscillates. Motion may be imparted to the bag in several ways, but the general effect is to create a sine wave along the fabric. As the fabric moves outward, accumulated dust on the surface moves with the fabric. When the fabric reaches the limit of its extension, the patches of dust have enough inertia to tear away from the fabric and descend to the hopper.

For small, single-compartment baghouses, a lever attached to the shaker mechanism may be operated manually at appropriate intervals, typically at the end of a shift. In multicompartment baghouses, a timer or a pressure sensor responding to system pressure drop initiates bag shaking automatically. The compartments operate in sequence so that one compartment at a time is cleaned. Forward gas flow to the compartment is stopped, dust is allowed to settle, residual gas flow stops, and the shaker mechanism is switched on for several seconds to a minute or more. The settling and shaking periods may be repeated, then the compartment is brought back online for filtering. Many large-scale shaker systems employ a small amount of reverse air during the shaker cycle to assist cleaning by deflating the bags.

Parameters that affect cleaning include the amplitude and frequency of the shaking motion and the tension of the mounted bag. The first two parameters are part of the baghouse design and generally are not changed easily. Typical values are about 4 Hz for frequency and 2 to 3 in. for amplitude (half-stroke).(1) The tension is set to about 2 lb/in. of bag circumference when bags are installed. Some installations allow easy adjustment of bag tension, while others require that the bag be loosened and reclamped to its attaching thimble.

The vigorous action of shaker systems tends to stress the bags and requires heavier and more durable fabrics. In the United States, woven fabrics are used almost exclusively(2) for shaker cleaning. European practice allows the use of felted fabrics at somewhat higher filtering velocities.

5.1.2.2 Reverse-air Cleaning

When glass fiber fabrics were introduced, a gentler means of cleaning the bags was needed to prevent premature degradation. Reverse-air cleaning was developed as a less intensive way to impart energy to the bags. In this method, gas flow to the bags is stopped in the compartment being cleaned, and a reverse flow of air is directed through the bags. This reversal of gas flow gently collapses the bags and dust is removed from the fabric surface by shear forces developed between the dust and fabric as the latter changes its contours. Another difference between reverse-air and shaker cleaning is the installation of sewn-in rings to prevent complete collapse of the bag, which may be greater than 30 ft long, during cleaning. Without these rings, collected dust tends to choke the bag as the fabric collapses in on itself. As with multicompartment shaker baghouses, the same cycle takes place in reverse-air baghouses of stopping forward gas flow and allowing dust to settle before cleaning action begins.

The source of reverse air is generally a separate fan capable of supplying air for one or two compartments at a gas-to-cloth ratio similar to that of the forward gas flow.

5.1.2.3 Pulse-jet Cleaning

This form of cleaning uses compressed air to force a burst of air down through the bag and expand it violently. As with shaker baghouses, the fabric reaches its extension limit and the dust separates from the bag. In pulse jets, however, gas flows are opposite in direction when compared with shaker or reverse-air baghouses. Bags are mounted on wire cages to prevent collapse while the dusty gas flows from outside the bag to the inside. Instead of attaching both ends of the bag to the baghouse structure, the bag and cage assembly generally is attached only at the top. The bottom end of the assembly tends to move in the turbulent gas flow and may contact other bags, which accelerates wear.

Although some pulse-jet baghouses are compartmented, most are not. Bags are cleaned by rows when a timer initiates the burst of cleaning air through a quick-opening valve. Usually 10% of the collector is pulsed at a time by zones. A pipe across each row of bags carries the compressed air. The pipe is pierced above each bag so that cleaning air exits directly downward into the bag. Some systems direct the air through a short venturi that is intended to entrain additional cleaning air. The pulse interrupts forward gas flow only for a few tenths of a second. However, the quick resumption of forward flow redeposits most of the dust back on the clean bag or on adjacent bags. An advantage of online pulse-jet cleaning is the reduction in baghouse size allowed by not having to build an extra compartment for offline cleaning. Pulse jets normally operate at two or more times the gas-to-cloth ratio of reverse-air baghouses.

5.1.3 Auxiliary Equipment

The typical auxiliary equipment associated with fabric filter systems is shown in Figure 5-1. Along with the fabric filter itself, a control system typically includes the following auxiliary equipment: a capture device (i.e., hood or direct exhaust connection); ductwork; dust removal equipment (screw conveyor, etc.); fans, motors, and starters; and a stack. In addition, spray chambers, mechanical collectors, and dilution air ports may be needed to precondition the gas before it reaches the fabric filter. Capture devices are usually hoods that exhaust pollutants into the ductwork or direct exhaust couplings attached to a process vessel. Hoods are more common, yet poorly

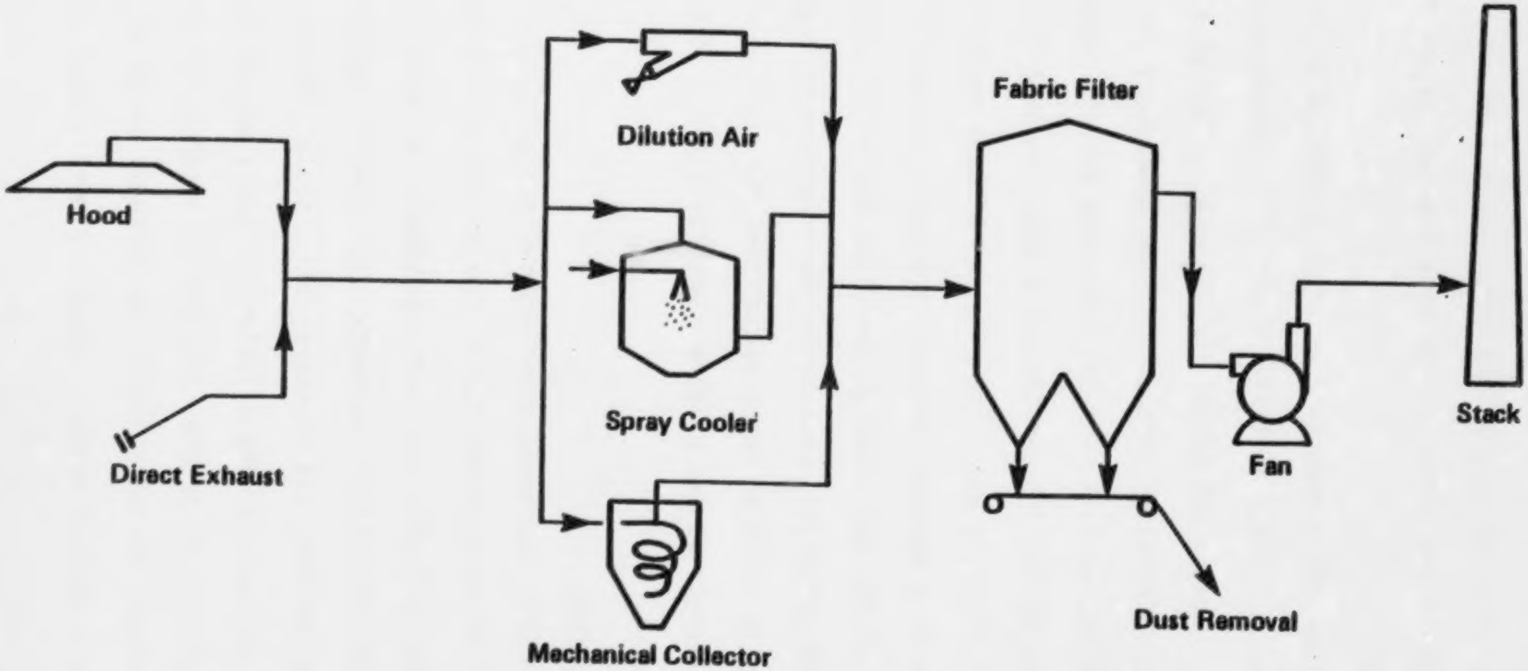
Hood
Direct Exhaust
Dilution Air
Spray Cooler
Mechanical Collector
Fabric Filter
Dust Removal
Fan
Stack

Figure 5-1. Typical alternative auxiliary equipment items used with fabric filter control systems.

designed hoods will allow pollutants to escape. Direct exhaust couplings are less common, requiring sweep air to be drawn through the process vessel, and may not be feasible in some processes. Ductwork provides a means of moving the exhaust stream to the control device. Spray chambers and dilution air ports are used to decrease the temperature of the pollutant stream to protect the filter fabric from excessive temperatures. When a substantial portion of the pollutant loading consists of relatively large particles, mechanical collectors such as cyclones are used to reduce the load on the fabric filter itself. The fans provide motive power for air movement and can be mounted before (pressure baghouse) or after (suction baghouse) the filter. A stack, when used, vents the cleaned stream to the atmosphere. Screw conveyors are often used to remove captured dust from the bottom of the hoppers. Air conveying systems and direct dumping into containers are also used.

5.1.4 Fabric Filtration Theory

The key to designing a baghouse is to determine the face velocity that produces the optimum balance between pressure drop (operating cost) and baghouse size (capital cost). Major factors that affect design face velocity (or gas-to-cloth ratio), discussed in Section 5.2, include particle and fabric characteristics and gas temperature. Although collection efficiency is another important measure of baghouse performance, it is generally assumed that a properly designed and well run baghouse will be highly efficient. Therefore, the design process focuses on the pressure drop. There are several contributions to the pressure drop across a baghouse including the pressure drop from the flow through the inlet and outlet ducts, from flow through the hopper regions, and from flow through the bags. The pressure drop through the baghouse (excluding the pressure drop across the bags) depends largely on the baghouse design and ranges from 1 to 2 inches of $H_2O^{(3)}$ in conventional designs and up to 3 inches of H_2O in designs having complicated gas flow paths. This loss can be kept to a minimum (i.e., 1 inch of H_2O or less) by investing in a flow modeling study of the proposed design. A study of this sort would cost on the order of $50,000 (in 1986). The pressure drop across the bags (also called the tubesheet pressure drop) can be as high as 10 inches of H_2O or more. The duct and hopper losses are constant and can be minimized effectively through proper design based on a knowledge of the flow through the baghouse. (Note: A procedure for estimating duct pressure losses is given in the "Ductwork" section of this Manual.) The tubesheet pressure drop is a complex function of the physical properties of the dust and fabric and the manner in which the baghouse is designed and operated.

Fabric filtration is inherently a batch process that has been adapted to continuous operation through clever engineering. One requirement for a continuously operating baghouse is that the dust collected on the bags must be removed periodically. Shaker and reverse-air baghouses are similar in the sense that they both normally use woven fabric bags, run at relatively low face velocities, and the filtration mechanism is cake filtration. That is, the fabric merely serves as a substrate for the formation of a dust cake that is the actual filtration medium. Pulse-jet baghouses generally use felt fabrics and run with a high face velocity (about double that of shaker or reverse-air baghouses). Some investigators feel that the felt fabric plays a much more active role in the filtration process. This distinction between cake filtration and fabric filtration has important implications for calculating the rate of pressure loss across the filter bags. The theoretical description of cake filtration is quite different from that for fabric filtration, and the design processes are quite different.

The general equations used to design a baghouse follow beginning with the reverse air/shake deflate type of baghouse.

5.1.4.1 Reverse Air/Shake Deflate Baghouses

The construction of a baghouse begins with a set of specifications including average pressure drop, total gas flow, and other requirements; a maximum pressure drop is always specified. Given these specifications, the designer must determine the maximum face velocity that can meet these requirements. The standard way to relate baghouse pressure drop to face velocity is given by the relation:

$$\Delta P(\theta) = S_{sys}(\theta) V_{ave} \qquad (5\text{-}1)$$

where:

$\Delta P(\theta)$ = the pressure drop across the filter, in. H_2O (a function of time, θ)

$S_{sys}(\theta)$ = system drag, in. H_2O/(ft/min) (a function of time)

V_{ave} = average (i.e., design) face velocity (ft/min) (essentially constant)

For a multicompartment baghouse, the system drag is determined as the sum of several parallel resistances representative of several compartments. For the typical case where the pressure drop through each compartment is the same, it can be shown that:

$$S_{sys}(\theta) = \left[\frac{1}{M} \sum_{i=1}^{M} \frac{1}{S_i(\theta)} \right]^{-1} = \frac{1}{\frac{1}{M} \sum\limits_{i=1}^{M} \frac{1}{S_i(\theta)}} = \frac{M}{\sum\limits_{i=1}^{M} \frac{1}{S_i(\theta)}} \qquad (5\text{-}2)$$

where

M = number of compartments in the baghouse

$S_i(\theta)$ = drag across compartment i

The compartment drag is a function of the amount of dust collected on the bags in that compartment. In general, the dust will be distributed in a very nonuniform manner. That is, there will be a variation of dust load from one bag to the next and within a given bag there will also be a variation of dust load from one area to another. For a sufficiently small area j within compartment i, it can be assumed that the drag is a linear function of dust load:

$$S_{i,j}(\theta) = S_e + K_2 W_{i,j}(\theta) \qquad (5\text{-}3)$$

where:

S_e = drag of a dust-free (freshly cleaned) filter bag

K_2 = dust cake flow resistance, [in. $H_2O/(ft/min)]/(lb/ft^2)$

$W_{i,j}(\theta)$ = dust mass per unit area of area j in compartment i

If there are N different areas of equal size within compartment i, each with a different drag $S_{i,j}$, then the total drag for compartment i can be computed in a manner analogous to equation (5-2):

$$S_i(\theta) = N/\Sigma[1/S_{i,j}(\theta)] \quad . \qquad (5\text{-}4)$$

The constants S_e and K_2 depend upon the fabric and the nature and size of the dust. The relationships between these constants and the dust and fabric properties are not understood well enough to permit accurate predictions and so must be determined empirically, either from prior experience with the dust/fabric combination or from laboratory measurements. The dust mass as a function of time is defined as:

$$W_i(\theta) = W_r + \int_0^{\theta} \{C_{in} V_i(\theta) d\theta\} \qquad (5\text{-}5)$$

where:

W_r = dust mass per unit area remaining on a "clean" bag

C_{in} = dust concentration in the inlet gas (gr/ft^3)

$V_i(\theta)$ = face velocity in compartment i

It is assumed that the inlet dust concentration and the filter area are constant. The face velocity through each compartment changes with time, starting at a maximum value just after cleaning and steadily decreasing as dust builds up on the bags. The individual compartment face velocities are related to the average face velocity by the expression:

$$V_{ave} = \Sigma\{V_i(\theta)A_i\}/\Sigma\{A_i\} \tag{5-6}$$

$$= \Sigma\{V_i\}/M \text{ (for M compartments with equal area)}$$

Equations (5-1) through (5-6) reveal that there is no explicit relationship between the design face velocity and the tubesheet pressure drop. On the contrary, the pressure drop that results from a given design can only be determined by the simultaneous solution of equations (5-1) through (5-5), with equation (5-6) as a constraint on that solution. This conclusion has several implications for the design process. The design requires an iterative procedure: one must begin with a known target for the average pressure drop, propose a baghouse design (number of compartments, length of filtration period, etc.), assume a face velocity that will yield that pressure drop, and solve the system of equations (5-1) through (5-6) to verify that the calculated pressure drop equals the target pressure drop. This procedure is repeated until the specified face velocity yields an average pressure drop (and maximum pressure drop, if applicable) that is sufficiently close to the design specification.

5.1.4.2 Pulse-Jet Baghouses

The distinction between pulse-jet baghouses and reverse-air and shake baghouses is basically the difference between cake filtration and composite dust/fabric filtration (noncake filtration). This distinction is more a matter of convenience than physics. In reality, pulse-jet baghouses have been designed to operate in a variety of modes. Some pulse jets remain online at all times and are cleaned frequently. Others are taken offline for cleaning at relatively long intervals. Obviously, if a compartment remains online long enough without being cleaned, then the filtration mechanism becomes that of cake filtration. A complete model of pulse-jet filtration therefore must account for the depth filtration occurring on a relatively clean pulse-jet filter, the cake filtration that inevitably results from prolonged periods online, and the transition period between the two regimes.

Besides the question of filtration mechanism, there is also the question of cleaning method. If a compartment is taken off-line for cleaning, then the dust that is removed from the bags will fall into the dust hopper before forward gas flow resumes. If a compartment is cleaned while online, then only a small fraction of the dust removed from the bag will fall to the hopper. The remainder of the dislodged dust will be redeposited (i.e., "recycled") on the bag by the forward gas flow. The redeposited dust layer has different pressure drop characteristics than the freshly deposited dust. The modeling work that has been done to date focuses on the online cleaning method. Dennis and Klemm(4) proposed the following model of drag across a pulse-jet filter:

$$S = S_e + (K_2)_c W_c + K_2 W_o \qquad (5\text{-}7)$$

where:

S = drag across the filter

The disadvantage of the model represented by equations (5-7) and (5-8) is that the constants, S_e, K_2, and W_c, cannot be predicted at this time. Consequently, correlations of laboratory data must be used to determine the value of $(PE)_{\Delta w}$. For the fabric-dust combination of Dacron felt and coal fly ash, Dennis and Klemm(4) developed an empirical relationship between $(PE)_{\Delta w}$, the face velocity, and the cleaning pulse pressure. This relationship (converted from metric to English units) was as follows:

$$(PE)_{\Delta w} = 6.08\ VP_j^{-0.65} \qquad (5\text{-}10)$$

where:

V = face velocity (ft/min)

P_j = pressure of the cleaning pulse (usually 60 to 100 psig; see Section 5.4.1.8)

It is not known how well the constants in equation (5-10) would fit the data for a different dust/fabric combination. Based on a limited amount of data, it appears that the power law form of equation (5-10) may be a valid model for $(PE)_{\Delta w}$. However, equation (5-10) can be used as a first approximation to estimate $(PE)_{\Delta w}$ for other fabric-dust combinations.

Another model that shows promise in the prediction of noncake filtration pressure drop is that of Leith and Ellenbecker(5) as modified by Koehler and Leith.(6) In this model, the tubesheet pressure drop is a function of the clean fabric drag, the system hardware, and the cleaning energy. Specifically:

$$\Delta P = (1/2)[P_s + K_1V_f - \sqrt{\{(P_s - K_1V_f)^2 - 4W_oK_2/K_3\}}] + K_vV_f^2 \qquad (5\text{-}11)$$

where:

P_s = maximum static pressure achieved in the bag during cleaning

K_1 = clean fabric resistance

V_f = face velocity

K_2 = dust deposit flow resistance

K_3 = bag cleaning efficiency coefficient

K_v = loss coefficient for the venturi at the inlet to the bag

Comparisons of laboratory data with pressure drops computed from equation (5-11)(5,6) are in close agreement for a variety of dust/fabric combinations. The disadvantage of equation (5-11) is that the constants K_1, K_2, and K_3 must be determined from laboratory measurements. The most difficult one to determine is the K_3 value, which can only be found by making measurements in a pilot-scale pulse-jet baghouse.

5.2 Design Procedures

5.2.1 Gas-to-Cloth Ratio

The gas-to-cloth ratio is difficult to estimate from first principles. However, shortcut methods of varying complexity allow rapid estimation. Descriptions of three methods of increasing difficulty follow. For shaker and reverse-air baghouses, the third method is best performed with publicly available computer model programs.

5.2.1.1 Gas-to-Cloth Ratio From Similar Applications

Net gas-to-cloth ratio is equal to the total actual volumetric flow rate in cubic feet per minute divided by the net cloth area in square feet. This ratio reduces to units of feet per minute and affects pressure drop and bag life. After a fabric has been selected, the gas-to-cloth ratio can be determined using Table 5-1. Column 1 shows the type of dust; column 2 shows the gas-to-cloth ratios for woven fabric; and column 3 shows gas-to-cloth ratios for felted fabrics. The net cloth area is determined by dividing the

S_e = drag of a just-cleaned filter

$(K_2)_c$ = specific dust resistance of the recycling dust

W_c = areal density of the recycling dust

K_2 = specific dust resistance of the freshly deposited dust

W_o = areal density of the freshly deposited dust

This model has the advantage that it can easily account for all three regimes of filtration in a pulse-jet baghouse, i.e., cake filtration, depth filtration, and filtration in the transition region. As in equations (5-1) to (5-6), the drag and areal densities are functions of time (θ). However, for a pulse-jet baghouse with online cleaning, the filtration velocity is relatively constant. The pressure drop can thus be expressed as the sum of a relatively constant term and a term that increases due to dust build-up:

$$\Delta P = (PE)_{\Delta w} + K_2 W_o V \tag{5-8}$$

where:

ΔP = pressure drop (in. H_2O)

V = filtration velocity (ft/min)

$$(PE)_{\Delta w} = [S_e + (K_2)_c W_c]\ V \tag{5-9}$$

Equation (5-8) describes the pressure drop behavior of an individual bag. To extend this single bag result to a multiple-bag compartment, equation (5-7) would be used to determine the individual bag drag and the total baghouse drag would then be computed as the sum of the parallel resistances as in equation (5-2). Pressure drop would then be calculated as in equation (5-1). It seems reasonable to extend this analysis to the case where the dust is distributed unevenly on the bag and then apply equation (5-7) to each area on the bag, followed by an equation analogous to (5-4) to compute the overall bag drag. The difficulty in doing this is that one must assume values for W_c for each different area to be modeled.

Table 5-1 Gas-to-Cloth Ratios[a](7)
$(ft^3/min)/(ft^2$ of cloth area)

Dust	Shaker/Woven Reverse-Air/Woven	Pulse Jet/Felt
Alumina	2.5	8
Asbestos	3.0	10
Bauxite	2.5	8
Carbon Black	1.5	5
Coal	2.5	8
Cocoa, Chocolate	2.8	12
Clay	2.5	9
Cement	2.0	8
Cosmetics	1.5	10
Enamel Frit	2.5	9
Feeds, Grain	3.5	14
Feldspar	2.2	9
Fertilizer	3.0	8
Flour	3.0	12
Fly Ash	2.5	5
Graphite	2.0	5
Gypsum	2.0	10
Iron Ore	3.0	11
Iron Oxide	2.5	7
Iron Sulfate	2.0	6
Lead Oxide	2.0	6
Leather Dust	3.5	12
Lime	2.5	10
Limestone	2.7	8
Mica	2.7	9
Paint Pigments	2.5	7
Paper	3.5	10
Plastics	2.5	7
Quartz	2.8	9
Rock Dust	3.0	9
Sand	2.5	10
Sawdust (Wood)	3.5	12
Silica	2.5	7
Slate	3.5	12
Soap, Detergents	2.0	5
Spices	2.7	10
Starch	3.0	8
Sugar	2.0	7
Talc	2.5	10
Tobacco	3.5	13
Zinc Oxide	2.0	5

[a] Generally safe design values--application requires consideration of particle size and grain loading.

gas-to-cloth ratio into the actual cubic feet per minute flow of the exhaust gas stream. For an intermittent-type baghouse that is shut down for cleaning, this is the total, or gross, cloth area. However, for continuously operated filters, the area must be increased to allow the shutting down of one or more compartments for cleaning. Table 5-2 provides a guide for adjusting the net area to the gross area, which determines the size of a continuously cleaned filter.

5.2.1.2 Gas-to-Cloth Ratio From Manufacturer's Methods

Manufacturers have developed nomographs and charts that allow rapid estimation of the gas-to-cloth ratio. Two examples are given below, one for shaker-cleaned baghouses and the other for pulse-jet cleaned baghouses.

For shaker baghouses, Table 5-3 gives a factor method for estimating the ratio. Ratios for several materials in different operations are presented, but are modified by factors for particle size and dust load. Directions and an example are included. Gas-to-cloth ratios for reverse-air baghouses would be about the same or a little more conservative compared to the Table 5-3 values.

For pulse-jet baghouses, another factor method(9) has been modified with equations to represent temperature, particle size, and dust load:

$$V = A \times B \times 2.647T^{-0.2335} \times (0.7471 + 0.0853 \ln D) \times 1.0873\ L^{-0.06021} \qquad (5\text{-}12)$$

where:

V = gas-to-cloth ratio, ft/min

A = material factor, from Table 5-4

B = application factor, from Table 5-4

T = temperature, °F (between 50 and 275)

Table 5-2 Approximate Guide to Estimate Gross Cloth Area(8)

Net Cloth Area (ft^2)	Gross Cloth Area (ft^2)
1-4,000	Multiply by 2
4,001-12,000	" 1.5
12,001-24,000	" 1.25
24,001-36,000	" 1.17
36,001-48,000	" 1.125
48,001-60,000	" 1.11
60,001-72,000	" 1.10
72,001-84,000	" 1.09
84,001-96,000	" 1.08
96,001-108,000	" 1.07
108,001-132,000	" 1.06
132,001-180,000	" 1.05
above 180,001	" 1.04

Table 5-3 Manufacturer's Factor Method for Estimating Gas-to-Cloth Ratios for Shaker Baghouses

A

4/1 RATIO		3/1 RATIO		2.5/1 RATIO		2/1 RATIO		1.5/1 RATIO	
MATERIAL	OPERATION	MATERIAL	OPERATION	MATERIAL	OPERATION	MATERIAL	OPERATION	MATERIAL	OPERATION
Cardboard	1	Asbestos	1, 7, 8	Alumina	2, 3, 4, 5, 6	Ammonium Phosphate Pert.	2, 3, 4, 5, 6, 7	Activated Charcoal	2, 4, 5, 6, 7
Feeds	2, 3, 4, 5, 6, 7	Aluminum Dust	1, 7, 8	Carbon Black	4, 5, 6, 7	Diatomaceous Earth	4, 5, 6, 7	Carbon Black	11, 14
Flour	2, 3, 4, 5, 6, 7	Fibrous Mat'l.	1, 4, 7, 8	Cement	3, 4, 5, 6, 7	Dry Petrochem.	2, 3, 4, 5, 6, 7, 14	Detergents	2, 4, 5, 6, 7
Grain	2, 3, 4, 5, 6, 7	Cellulose Mat'l.	1, 4, 7, 8	Coke	2, 3, 5, 6	Dyes	2, 3, 4, 5, 6, 7	Metal Fumes, Oxides and other Solid Dispersed Products	10, 11
Leather Dust	1, 7, 8	Gypsum	1, 3, 5, 6, 7	Ceramic Pigm.	4, 5, 6, 7	Fly Ash	10		
Tobacco	1, 4, 6, 7	Lime (Hydrated)	2, 4, 6, 7	Clay & Brick Dust	2, 4, 6, 12	Metal Powders	2, 3, 4, 5, 6, 7, 14		
Supply Air	13	Perlite	2, 4, 5, 6	Coal	2, 3, 6, 7, 12	Plastics	2, 3, 4, 5, 6, 7, 14		
Wood, Dust, Chips	1, 6, 7	Rubber Chem.	4, 5, 6, 7, 8	Kaolin	4, 5, 7	Resins	2, 3, 4, 5, 6, 7, 14		
		Salt	2, 3, 4, 5, 6, 7	Limestone	2, 3, 4, 5, 6, 7	Silicates	2, 3, 4, 5, 6, 7, 14		
		Sand *	4, 5, 6, 7, 9, 15	Rock, Ore Dust	2, 3, 4, 5, 6, 7	Starch	6, 7		
		Iron Scale	1, 7, 8	Silica	2, 3, 4, 5, 6, 7	Soaps	3, 4, 5, 6, 7		
		Soda Ash	4, 6, 7	Sugar	3, 4, 5, 6, 7				
		Talc	3, 4, 5, 6, 7						
		Machining Operation	1, 8						

CUTTING - 1
CRUSHING - 2
PULVERIZING - 3
MIXING - 4
SCREENING - 5
STORAGE - 6
CONVEYING - 7
GRINDING - 8
SHAKEOUT - 9
FURNACE FUME - 10
REACTION FUME - 11
DUMPING - 12
INTAKE CLEANING - 13
PROCESS - 14
BLASTING - 15

B

FINENESS FACTOR

MICRON SIZE	FACTOR
> 100	1.2
50 - 100	1.1
10 - 50 *	1.0
3 - 10	.9
1 - 3	.8
<1	.7

C

DUST LOAD FACTOR

Loading GR. CU. FT.	Factor
1 - 3	1.2
4 - 8	1.0
9 - 17	.95
18 - 40	.90
> 40	.85

This information constitutes a guide for commonly encountered situations and should not be considered a "hard-and-fast" rule. Air-to-cloth ratios are dependent on dust loading, size distribution, particle shape and "cohesiveness" of the deposited dust. These conditions must be evaluated for each application. The longer the interval between bag cleaning the lower the air-to-cloth ratio must be. Finely-divided, uniformly sized particles generally form more dense filter cakes and require lower air-to-cloth ratios than when larger particles are interspersed with the fines. Sticky, oily particles, regardless of shape or size, form dense filter cakes and require lower air-to-cloth ratios.

EXAMPLE: Foundry shakeout unit handling 26000 CFM and collecting 3500 #/ hr. of sand. The particle distribution shows 90% greater than 10 microns. The air is to exhaust to room in winter, to atmosphere in summer.

$$3500\ \#/_{HR} \div 60\ \frac{MIN.}{HR} \div 26000\ \frac{CU.\ FT.}{MIN.} \times 7000\ ^{GR}/_{\#} = 15.7\ \frac{GR}{CU.\ FT.}$$

*Chart A = 3/1 ratio, Chart B = Factor 1.0, Chart C = .95; 3 x .1 x .95 = 2.9 air to cloth ratio. 26000 ÷ 2.9 = 9,000 sq. ft.

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Table 5-4 Factors for Pulse-Jet Gas-to-Cloth Ratios(9)

A. Material Factor

15[a]	12	10	9.0	6.0[b]
Cake mix	Asbestos	Alumina	Ammonium phosphate-fertilizer	Activated carbon
Cardboard dust	Buffing dust	Aspirin	Cake	Carbon black (molecular)
Cocoa	Fibrous and cellulosic material	Carbon black (finished)	Diatomaceous earth	Detergents
Feeds	Foundry shakeout	Cement	Dry petrochemicals	Fumes and other dispersed products direct from reactions
Flour	Gypsum	Ceramic pigments	Dyes	Powdered milk
Grain	Lime (hydrated)	Clay and brick dusts	Fly ash	Soaps
Leather dust	Perlite	Coal	Metal powder	
Sawdust	Rubber chemicals	Fluorspar	Metal oxides	
Tobacco	Salt	Gum, natural	Pigments, metallic and synthetic	
	Sand	Kaolin	Plastics	
	Sandblast dust	Limestone	Resins	
	Soda ash	Perchlorates	Silicates	
	Talc	Rock dust, ores and minerals	Starch	
		Silica	Stearates	
		Sorbic acid	Tannic acid	
		Sugar		

B. Application Factor

Nuisance Venting Relief of transfer points, conveyors, packing stations, etc.	1.0
Product Collection Air conveying-venting mills flash driers classifiers, etc.	0.9
Process Gas Filtration Spray driers, kilns, reactors, etc.	0.8

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[a] In general physically and chemically stable materials.

[b] Also includes those solids that are unstable in their physical or chemical state due to hygroscopic nature, sublimation, and/or polymerization.

D = mass mean diameter of particle, μm (between 3 and 100)

L = inlet dust loading, gr/ft^3 (between 0.05 and 100)

For temperatures below 50 °F, use T = 50 but expect decreased accuracy; for temperatures above 275 °F, use T = 275. For particle mass mean diameters less than 3 μm, the value of D is 0.8, and for diameters greater than 100 μm, D is 1.2. For dust loading less than 0.05 gr/ft^3, use L = 0.05; for dust loading above 100 gr/ft^3, use L = 100.

5.2.1.3 Gas-to-Cloth Ratio From Theoretical/Empirical Equations

5.2.1.3.1 Shaker and reverse-air baghouses--The system described by equations (5-1) through (5-6) is complicated; however, numerical methods can be used to obtain an accurate solution. A critical weakness in baghouse modeling that has yet to be overcome is the lack of a fundamental description of the bag cleaning process. That is, to solve equations (5-1) through (5-6), the value of W_r (the dust load after cleaning) must be known. Clearly, there must be a relationship between the amount and type of cleaning energy and the degree of dust removal from a bag. Dennis et al.(10) have developed correlations for the removal of coal fly ash from woven fiberglass bags by shaker cleaning and by reverse air cleaning. These correlations have been incorporated into a computer program that generates the solution to the above system of equations.(10,11,12) If one were to apply the correlations developed with coal ash and woven glass fabrics to other dust/fabric combinations, the accuracy of the results would depend on how closely that dust/fabric combination mimicked the coal ash/woven glass fabric system. Physical factors that affect the correlation include the particle size distribution, adhesion properties of the dust and fabric, and fabric weave, as well as cleaning energy. More research is needed in this area of fabric filtration.

The rigorous design of a baghouse thus involves several steps. First, the design goal for average pressure drop (and maximum pressure drop, if necessary) must be specified along with total gas flow rate and other parameters, such as S_e and K_2 (obtained either from field or laboratory measurements). Second, a face velocity is assumed and the number of compartments in the baghouse is computed based on the total gas flow, face velocity, bag size, and number of bags per compartment (typical compartments in the U.S. electric utility industry use bags 1 ft in diameter by 30 ft long with 400 bags per compartment). Standard practice is to design a baghouse to meet the specified pressure drop when one compartment is off-line for maintenance and a second compartment off-line for cleaning. The third step is to specify the operating characteristics of the baghouse (i.e., filtration period, cleaning period, and cleaning mechanism). Fourth, the designer must specify the cleaning efficiency so that the residual dust load can be estimated. Finally, the specified baghouse design is used to establish the details for equations (5-1) through (5-6), which are then solved numerically to establish the pressure drop as a function of time. The average pressure drop is then computed by integrating the instantaneous pressure drop over the filtration cycle and dividing by the cycle time. If the computed average is higher than the design specification, then the face velocity must be reduced and the procedure repeated. If the computed average pressure drop is significantly lower than the design specification, then the proposed baghouse was oversized and should be made smaller by increasing the face velocity and repeating the procedure. When the computed average pressure drop comes sufficiently close to the assumed specified value, then the design has been determined. A complete description of the modeling process can be found in the reports by Dennis et al.(10,12) A critique on the accuracy of the model is presented by Viner et al.(13)

5.2.1.3.2 Pulse-jet baghouses--The overall process of designing a pulse jet baghouse is actually simpler than that required for a reverse-air or shaker baghouse if the baghouse remains online for cleaning. The first step is to specify what the desired average tubesheet pressure drop should be. Second, the operating characteristics of the baghouse must be established (e.g., online time, cleaning energy). Third, the designer must obtain values for the coefficients in either equation (5-10) or equation (5-11) from field, pilot plant, or laboratory measurements. Fourth, a value is estimated for the face velocity and the appropriate equation [(8) or (11)] is solved for the pressure drop as a function of time for the duration of the filtration cycle. This information is used to calculate the cycle average pressure drop. If the calculated pressure drop matches the specified pressure drop, then the procedure is finished. If not, then the designer must adjust the face velocity and repeat the procedure.

5.2.2 Pressure Drop

Pressure drop for the bags can be calculated rigorously from the equations given in the preceding section if values for the various parameters are known. Frequently they are not known. For quick estimation, a maximum pressure drop of 5- to 10-in. H_2O across the baghouse and 10- to 20-in H_2O across the entire system can be assumed if it contains much ductwork.

A comparable form of equations (5-1) and (5-3) that may be used for pressure drop across the fabric in a shaker or reverse-air baghouse is:

$$\Delta P = S_e V + K_2 C_i V^2 \theta \qquad (5\text{-}13)$$

where:

ΔP = pressure drop (in. H_2O)

S_e = effective residual drag of the fabric [in. H_2O/(ft/min)]

V = superficial face velocity or gas-to-cloth ratio (ft/min)

K_2 = specific resistance coefficient of the dust [in. H_2O/(ft/min)]/(lb/ft^2)

C_i = inlet dust concentration (lb/ft^3)

θ = filtration time, min

Although there is much variability, values for S_e may range from about 0.2 to 2 in. H_2O/(ft/min) and for K_2 from 1 or 2 to 30 or 40 [in. H_2O/(ft/min)]/lb/ft. Typical values for coal fly ash are 1 to 4. Inlet concentrations vary from less than 0.05 gr/ft^3 to more than 100 gr/ft^3, but a more nearly typical range is from 0.5 to 10 gr/ft^3. Filtration times may range from 20 minutes to 8 hours for continuous duty reverse-air and shaker baghouses, but 30 minutes to 4 hours is more frequently found. Filtration times for pulse-jet baghouses range from 2 to 60 minutes, but 5 to 20 minutes is typical. For pulse-jet baghouses, use equations (5-8) and (5-10) to estimate ΔP, after substituting C_iVt for W_o and $(PE)_{\Delta w}$ for S_e V.

5.2.3 Particle Characteristics

Particle size distribution and adhesiveness are the most important particle properties that affect design procedures. Smaller particle sizes can form a denser cake, which increases pressure drop. As shown in Table 5-3 and equation (5-12), the effect of decreasing average particle size is a lower applicable gas-to-cloth ratio.

Sticky particles may require installing equipment that injects a precoating material onto the bag surface, which acts as a buffer that traps the particles and prevents them from blinding or permanently plugging the fabric pores.

5.2.4 Gas Stream Characteristics

Moisture and corrosives content are the major gas stream characteristics requiring design consideration. The baghouse and associated ductwork should be insulated and possibly heated to avoid condensation. Both the structural and fabric components must be considered, as either may be damaged. Where structural corrosion is likely, stainless steel substitution for mild steel may be required, provided that chlorides are not present. (Most austenitic stainless steels are susceptible to chloride corrosion.)

5.2.4.1 Temperature

The temperature of the pollutant stream to be cleaned must be above and remain above the dew point of any condensables in the stream. If the temperature is high and it can be lowered without approaching the dew point, spray coolers or dilution air can be used to drop the temperature so that temperature limits of the fabric will not be exceeded. The additional cost of a precooler will have to be weighed against the higher cost of bags with greater temperature resistance. The use of dilution air to cool the stream also constitutes a tradeoff between a less expensive fabric and a larger filter necessary to accommodate the additional volume of the dilution air. Generally, precooling would not be necessary if fabric that will handle the temperature and the chemical action of the pollutant stream is available. (Costs for spray chambers, quenchers, and other precoolers are found in the "Precoolers" section of the Manual.) Table 5-5 lists several of the fabrics in current use and provides information on temperature limits and chemical resistance. The column labeled "Flex Abrasion" indicates the fabric's suitability for cleaning by mechanical shakers.

Table 5-5 Properties of Leading Fabric Materials(14)

Fabric	Temp, °F[a]	Acid Resistance	Alkali Resistance	Flex Abrasion
Cotton	180	Poor	Very good	Very good
Creslan[b]	250	Good in mineral acids	Good in weak alkali	Good to very good
Dacron[c]	275	Good in most mineral acids; disolves partially in concentrated H_2SO_4	Good in weak alkali; fair in strong alkali	Very good
Dynel[c]	160	Little effect even at high concentration	Little effect even in high concentration	Fair to good
Fiberglas[d]	500	Fair to good	Fair to good	Fair
Filtron[e]	270	Good to excellent	Good	Good to Very good
Gore-Tex[f]	Depends on backing	Depends on backing	Depends on backing	Fair
Nomex[c]	375	Fair	Excellent at low temperature	Excellent
Nylon[c]	200	Fair	Excellent	Excellent
Orlon[c]	260	Good to excellent in mineral acids	Fair to good in weak alkali	Good
Polypropylene	200	Excellent	Excellent	Excellent
Teflon[c]	450	Inert except to fluorine	Inert except to trifluoride, chlorine, and molten alkaline metals	Fair
Wool	200	Very good	Poor	Fair to good

[a] Maximum continuous operating temperatures recommended by the Industrial Gas Cleaning Institute.

[b] American Cyanamid registered trademark.

[c] Du Pont registered trademark.

[d] Owens-Corning Fiberglas registered trademark.

[e] W. W. Criswell Div. of Wheelabrator-Fry, Inc., trade name.

[f] W. L. Gore and Co., registered trademark.

5.2.4.2 Pressure

Standard fabric filters can be used in pressure or vacuum service but only within the range of about ±25 inches of water gauge. Because of the sheet metal construction of the house, they are not generally suited for more severe service. However, for special applications, high-pressure shells can be built.

5.2.5 Pressure or Suction Housings

The location of the baghouse with respect to the fan in the gas stream affects the capital cost. A suction-type baghouse, with the fan located on the downstream side of the unit, must withstand high negative pressures and therefore must be more heavily constructed and reinforced than a baghouse located downstream of the fan (pressure baghouse). The negative pressure in the suction baghouse can result in outside air infiltration, which can result in condensation, corrosion, or even explosions if combustible gases are being handled. In the case of toxic gases, this inward leakage can have an advantage over the pressure-type baghouse, where leakage is outward. The main advantage of the suction baghouse is that the fan handling the process stream is located at the clean-gas side of the baghouse. This reduces the wear and abrasion on the fan and permits the use of more efficient fans (backward-curved blade design). However, because for some designs the exhaust gases from each compartment are combined in the outlet manifold to the fan, locating compartments with leaking bags may be difficult and adds to maintenance costs.

Pressure-type baghouses are generally less expensive because the housing must only withstand the differential pressure across the fabric. In some designs the baghouse has no external housing. Maintenance also is reduced because the compartments can be entered and leaking bags can be observed while the compartment is in service. With a pressure baghouse, the housing acts as the stack to contain the fumes with the subsequent discharge at the roof of the structure, which makes it easier to locate leaking bags. The main disadvantage of the pressure-type baghouse is that the fan is exposed to the dirty gases where abrasion and wear on the fan blades may become a problem. Also, some applications require a stack for dispersion of gaseous pollutants, negating some of the construction economics.

5.2.6 Standard or Custom Construction

The design and construction of baghouses are separated into two groups, standard and custom,(8) which are further separated into low, medium, and high capacity. Standard baghouses are predesigned and factory built as complete off-the-shelf units that are shop-assembled and bagged for low-capacity units (hundreds to thousands of acfm throughput). Medium-capacity units (thousands to less than 100,000 acfm) have standard designs, are shop-assembled, may or may not be bagged, and have separate bag compartment and hopper sections. High-capacity baghouses (larger than 50,000 or 100,000 acfm) can be designed as shippable modules requiring only moderate field assembly. These modules may have bags installed and can be shipped by truck or rail. Upon arrival, they can be operated singly or combined to form units for larger-capacity applications. Because they are preassembled, field labor for installation is less costly.

The custom baghouse, also high capacity, is designed for a specific application and is usually built to the specifications prescribed by the customer. Generally, these units are much larger than standard baghouses. For example, many are used on power plants. The cost of the custom baghouse is much higher per square foot of fabric because it is not an off-the-shelf item and requires special setups for manufacture and expensive field labor for assembly upon arrival. The advantages of the custom baghouse are many and are usually directed towards ease of maintenance, accessibility, and other customer preferences. In some very small baghouses, a complete set of bags must be replaced in a compartment at one time because of the difficulty in locating and replacing single leaking bags, whereas in custom baghouses, single bags are accessible and can be replaced one at a time as leaks develop.

5.2.7 Filter Media

The type of filter material used in baghouses is dependent on the specific application in terms of chemical composition of the gas, operating temperature, dust loading, and the physical and chemical characteristics of the particulate. A variety of fabrics, either felted or woven, is available and the selection of a specific material, weave, finish, or weight is based primarily on past experience. The type of yarn (filament, spun, or staple), the yarn diameter, and twist are also factors in the selection of suitable fabrics for a specific application. For some difficult applications, Gore-Tex, a polytetrafluoroethylene (PTFE) membrane laminated to a substrate fabric (felt or woven) or altered-surface fabrics may be used. Because of the violent agitation of mechanical shakers, spun or heavy weight staple yarn fabrics are commonly used with this type of cleaning, while lighter weight filament yarn fabrics are used with reverse-air cleaning.

The type of material will limit the maximum operating gas temperature for the baghouse. Cotton fabric has the least resistance to high temperatures (about 180 °F), while fiberglass has the most (about 500 °F). The temperature of the exhaust-gas stream must be well above the dew point of any of its contained condensables as liquid particles will usually plug the fabric pores quickly. However, the temperature must be below the maximum limit of the fabric in the bags. These maximum limits are given in Table 5-5.

5.3 Estimating Total Capital Investment

Total capital investment includes costs for the baghouse structure, the initial complement of bags, auxiliary equipment, and the usual direct and indirect costs associated with installing or erecting new structures. These costs are described below.

5.3.1 Equipment Cost

5.3.1.1 Bare Baghouse Costs

Six types of baghouses will be considered:

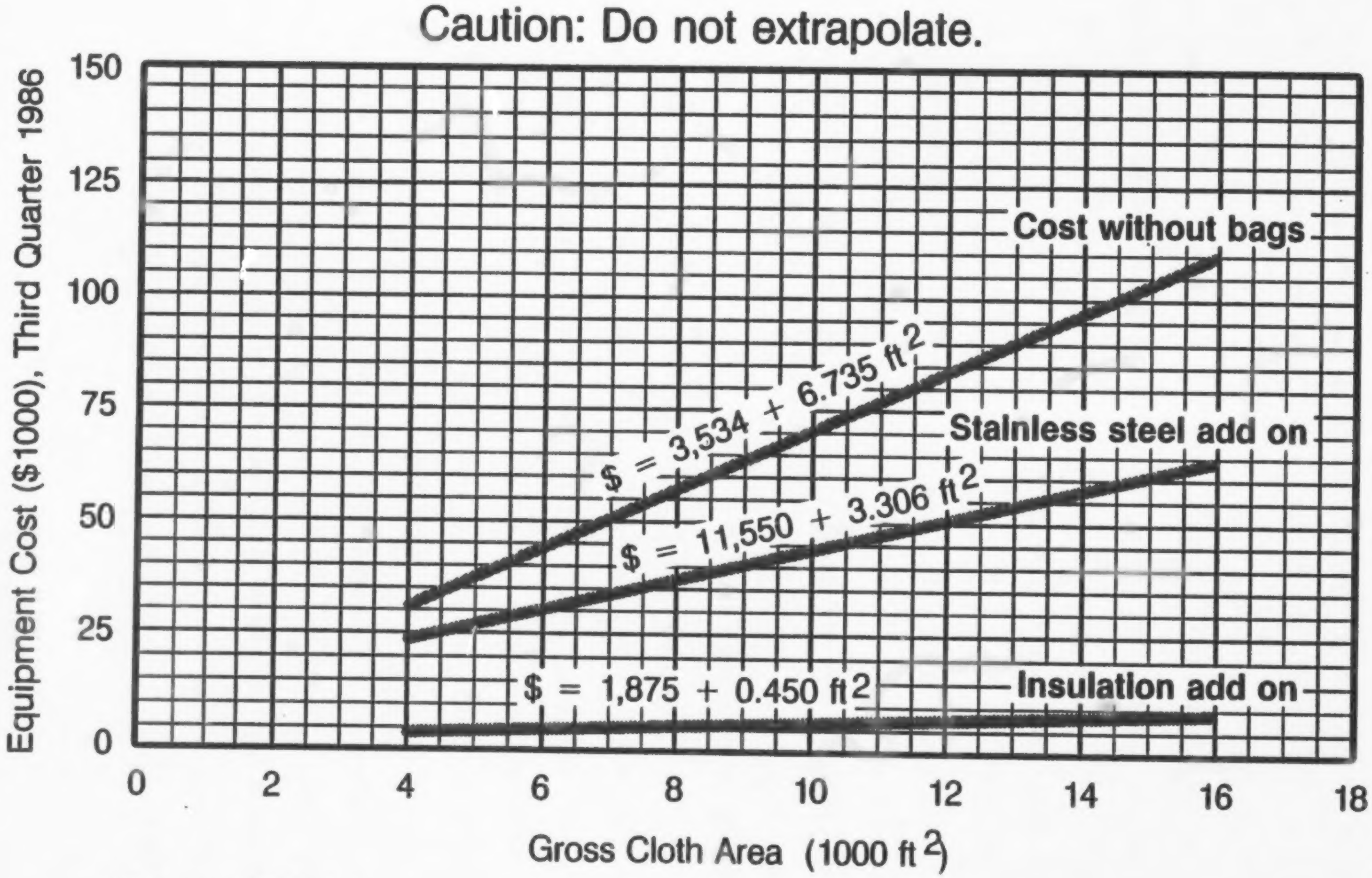
	Preassembled units	
Intermittent	Shaker	Figure 5-2
Continuous	Shaker	Figure 5-3
Continuous	Pulse-jet (common housing)	Figure 5-4
Continuous	Pulse-jet (modular)	Figure 5-5
Continuous	Reverse-air	Figure 5-6
	Field-assembled units	
Continuous	Any method	Figure 5-7

Each figure gives costs for the filter without bags and additional costs for stainless steel construction and for insulation. All curves are based on a number of actual quotes. A least squares line has been fitted to the quotes and the line's equation is given. However, extrapolation should not be used. The reader should not be surprised if he obtains quotes that differ from these curves by as much as ±25%. Significant savings can be obtained by soliciting multiple quotes. All units include inlet and exhaust manifolds, supports, platforms, handrails, and hopper discharge devices. The indicated prices are flange to flange. Note that the scales on both axes change from one figure to another to accommodate the differing gas flow ranges over which the various types of baghouses operate.

The 304 stainless steel add-on cost is used when such construction is necessary to prevent the exhaust gas stream from corroding the interior of the baghouse. Stainless steel is substituted for all metal surfaces that are in contact with the exhaust gas stream.

Insulation costs are for 3 inches of shop-installed glass fiber encased in a metal skin. One exception is the custom baghouse, which has field-installed insulation. Costs for insulation include only the flange-to-flange baghouse structure on the outside of all areas in contact with the exhaust gas stream. Insulation for ductwork, fan casings, and stacks must be calculated separately as discussed later.

The first baghouse type is the intermittent service baghouse cleaned by a mechanical shaker. This baghouse is shut down and cleaned at convenient times, such as the end of the shift or end of the day. Although few units are sold, they are applicable for operations that require infrequent cleaning. Figure 5-2 presents the unit cost with price in dollars plotted against the gross square feet of cloth required.[15] Because intermittent service baghouses do not require an extra compartment for cleaning, gross and net fabric areas are the same. The plot is linear because baghouses are made up of modular compartments and thus have little economy of scale. Because of the modular construction, the price line should not be extrapolated downward. Costs for both types of shaker baghouse include the shaker mechanism.



Source: ETS, Inc.

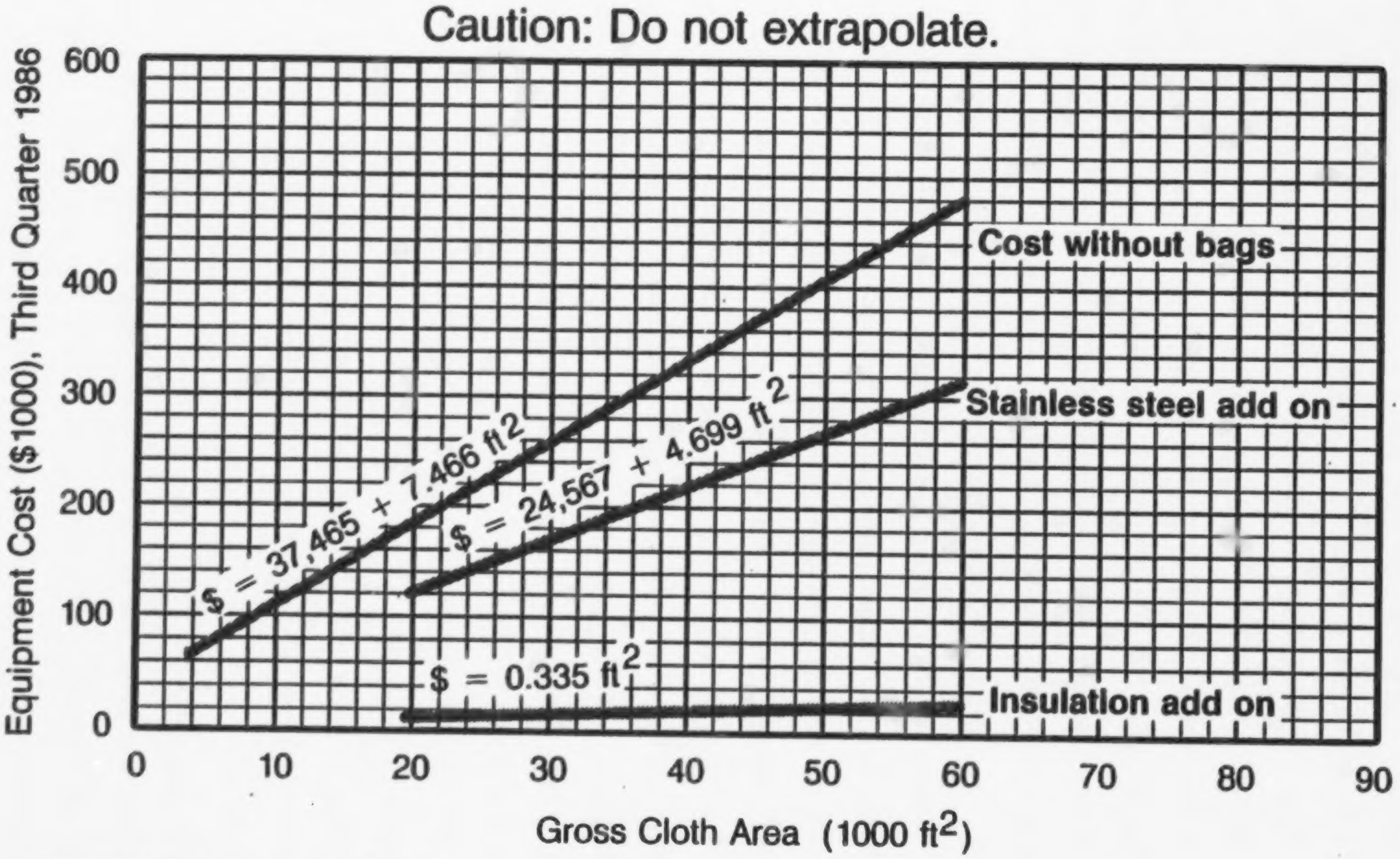
Figure 5-2. Equipment costs for intermittent shaker filters.

Figure 5-3 presents the same costs for a continuously operated baghouse cleaned by mechanical shaker.[15,16] Again, price is plotted against the gross cloth area in square feet. As in Figure 5-2, the units are modular in construction. Costs for these units, on a square foot basis, are higher because of increased complexity and generally heavier construction.

The third and fourth types are common-housing pulse jets and modular pulse jets. The latter are constructed of separate modules that may be arranged for offline cleaning, and the former have all bags within one housing. The costs for these units are shown in Figures 5-4 and 5-5, respectively.[15] The cleaning system compressor is not included. Note that in the single-unit (common-housing) pulse jet, for the range shown, the height and width of the unit are constant and the length increases; thus, for a different reason than that for the modular units discussed above, the cost increases linearly with size. Because the common housing is relatively inexpensive, the stainless steel add-on is proportionately higher than for modular units. Added material costs and setup and labor charges associated with the less workable stainless steel account for most of the added expense. Figure 5-6 shows the costs for the reverse-air baghouses.[15] The construction is modular and the reverse-air fan is included. The final type is the custom baghouse which, because of its large size, must be field assembled. It is often used on power plants, steel mills, or other applications too large for the factory-assembled baghouses. Prices for these units are shown in Figure 5-7.[15]

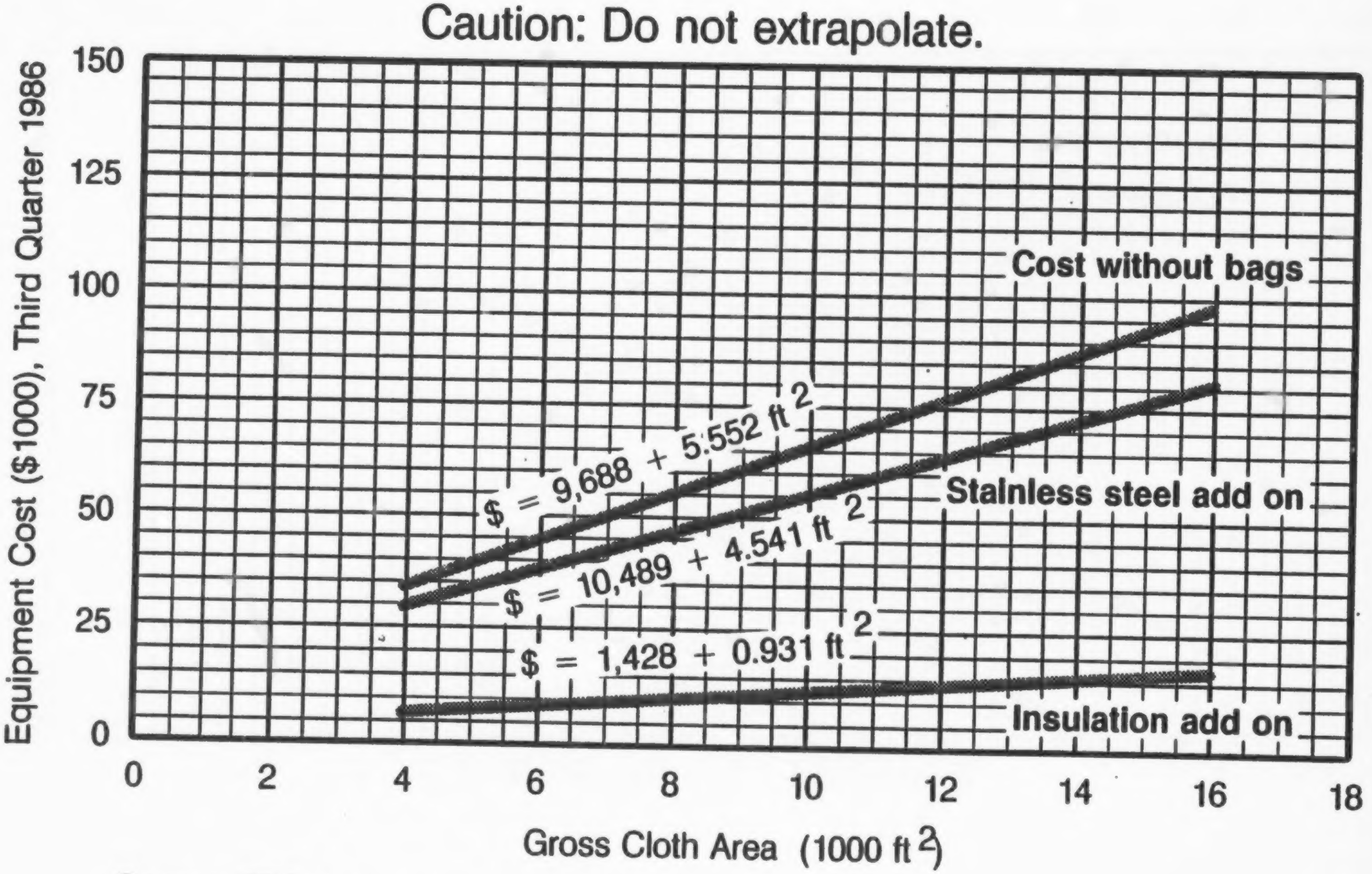
5.3.1.2 Bag Costs

Table 5-6 gives the price per square foot of bags by type of fabric and by type of cleaning system used. The prices represent about a 10% range. In calculating the cost, the gross area as determined from Table 5-2 should be



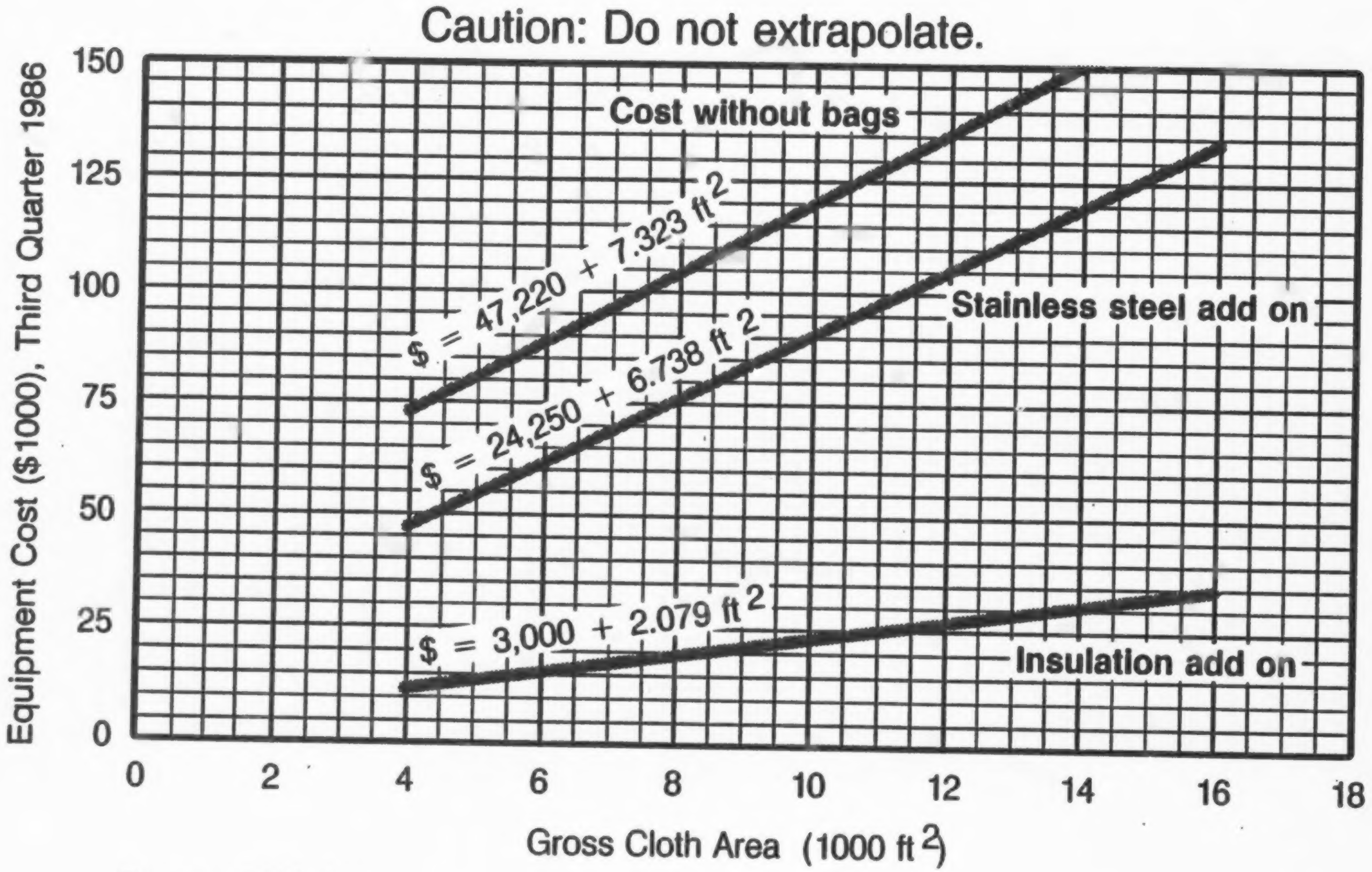
Source: ETS, Inc.; Fuller Co.

Figure 5-3. Equipment costs for continuous shaker filters.



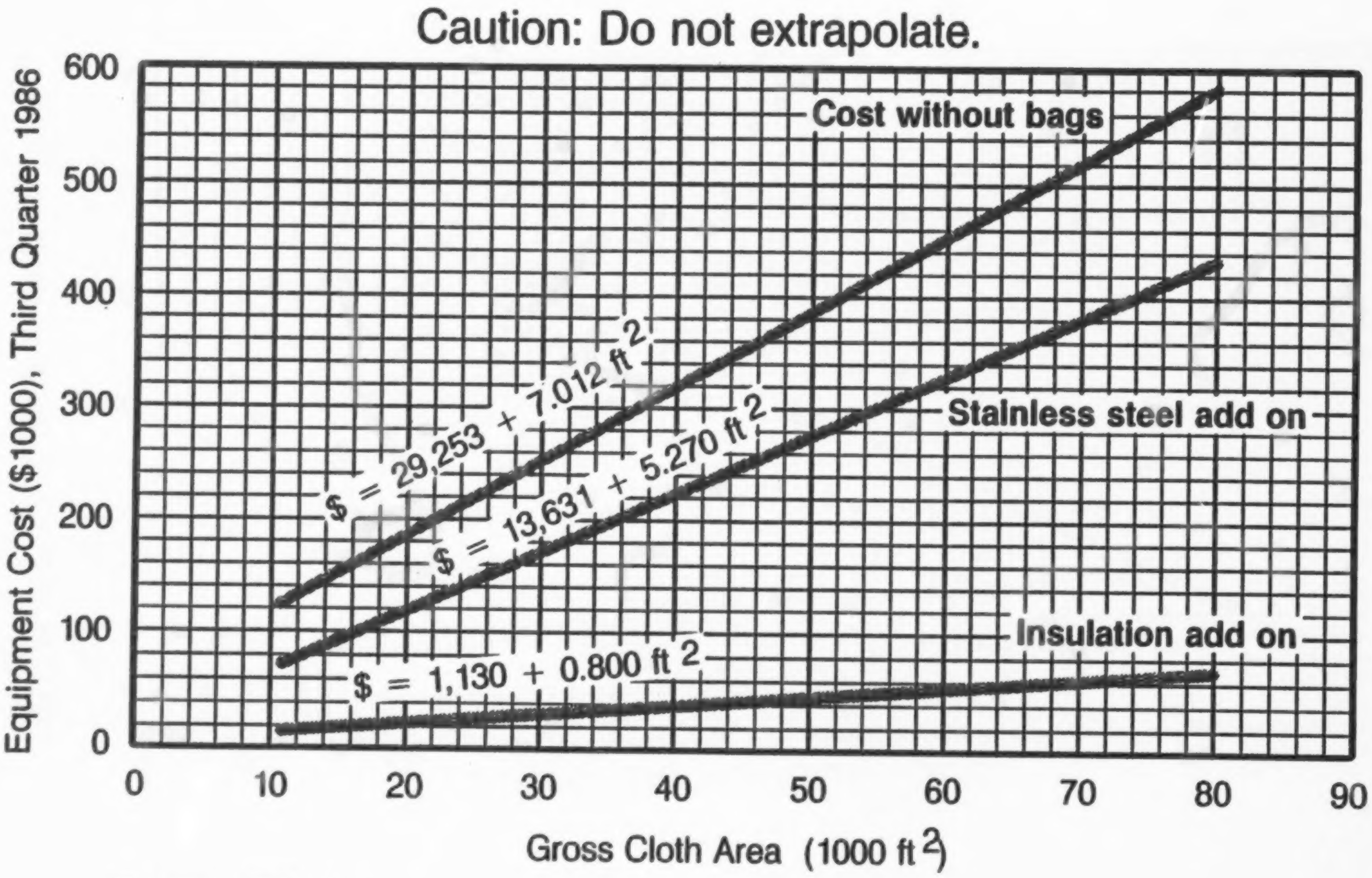
Source: ETS, Inc.

Figure 5-4. Equipment costs for pulse-jet filters (common housing).



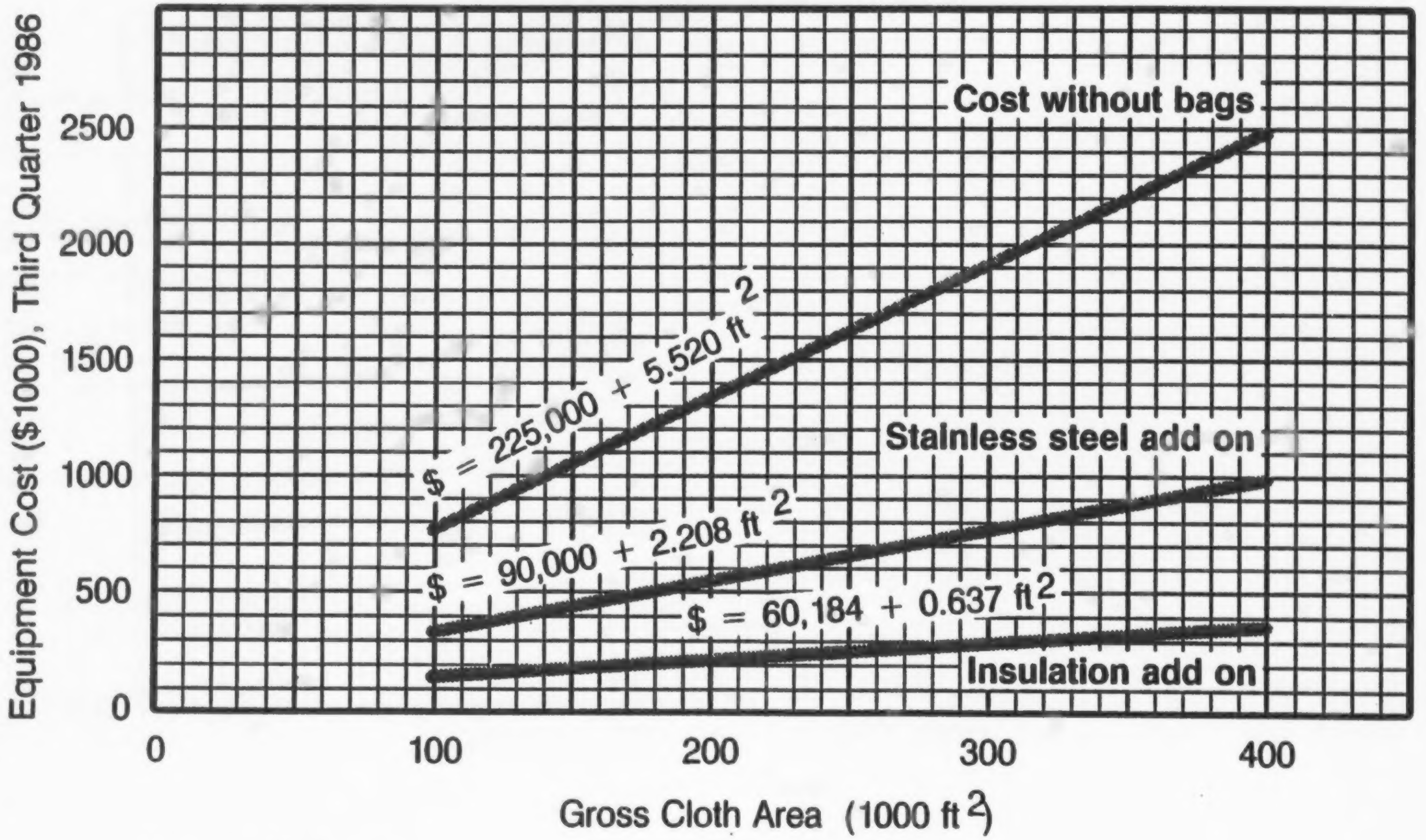
Source: ETS, Inc.

Figure 5-5. Equipment costs for pulse-jet filters (modular).



Source: ETS, Inc.

Figure 5-6. Equipment costs for reverse-air filters.



Caution: Do not extrapolate.

Source: ETS, Inc.

Figure 5-7 Equipment costs for custom-built filters.

Table 5-6 Bag Prices
(3rd quarter 1986 $\$/ft^2$)

Type of Cleaning	Bag Diameter (inches)	Type of Material[a]						
		PE	PP	NO	HA	FG	CO	TF
Pulse jet, TR[b]	4-1/2 to 5-1/8	0.59	0.61	1.88	0.92	1.29	NA	9.05
	6 to 8	0.43	0.44	1.56	0.71	1.08	NA	6.80
Pulse jet, BBR	4-1/2 to 5-1/8	0.37	0.40	1.37	0.66	1.24	NA	8.78
	6 to 8	0.32	0.33	1.18	0.58	0.95	NA	6.71
Shaker								
Strap top	5	0.45	0.48	1.28	0.75	NA	0.44	NA
Loop top	5	0.43	0.45	1.17	0.66	NA	0.39	NA
Reverse air with rings	8	0.46	NA	1.72	NA	0.99	NA	NA
	11-1/2	0.47	NA	1.69	NA	0.76	NA	NA
Reverse air w/o rings[c]	8	0.32	NA	1.20	NA	0.69	NA	NA
	11-1/2	0.32	NA	1.16	NA	0.53	NA	NA

NA = Not applicable.

[a] Materials:

PE = 16-oz polyester
PP = 16-oz polypropylene
NO = 14-oz nomex
HA = 16-oz homopolymer acrylic
FG = 16-oz fiberglass with 10% Teflon
CO = 9-oz cotton
TF = 22-oz Teflon felt

[b] Bag removal methods:

TR = Top bag removal (snap in)
BBR = Bottom bag removal

[c] Identified as reverse-air bags, but used in low pressure pulse applications.

NOTE: For pulse-jet baghouses, all bags are felts except for the fiberglass, which is woven. For bottom access pulse jets, the cage price for one cage can be calculated from the single-bag fabric area using:

	Mild steel cage	Stainless steel cage
In 50 cage lots	$\$ = 4.941 + 0.163\ ft^2$	$\$ = 23.335 + 0.280\ ft^2$
In 100 cage lots	$\$ = 4.441 + 0.163\ ft^2$	$\$ = 21.791 + 0.263\ ft^2$
In 500 cage lots	$\$ = 3.941 + 0.163\ ft^2$	$\$ = 20.564 + 0.248\ ft^2$

These costs apply to 4-1/2-in. or 5-5/8-in diameter, 8-ft and 10-ft cages made of 11 gauge mild steel and having 10 vertical wires and "Roll Band" tops. For flanged tops, add $1 per cage. If flow control venturis are used (as they are in about half of the pulse-jet manufacturers' designs), add $5 per cage.

For shakers and reverse air baghouses, all bags are woven. All prices are for finished bags, and prices can vary from one supplier to another. For Gore-Tex bag prices, multiply base fabric price by factors of 3 to 4.5.

Source: ETS, Inc.(15)

used. Gore-Tex fabric costs are a combination of the base fabric cost and a premium for the PTFE laminate and its application. As fiber market conditions change, the costs of fabrics relative to each other also change. The bag prices are based on typical fabric weights, in ounces/square yard, for the fabric being priced. Sewn-in snap rings are included in the price, but other mounting hardware, such as clamps or cages, is an added cost.

5.3.1.3 Auxiliary Equipment

The auxiliary equipment depicted in Figure 5-1 is discussed elsewhere in the Manual. Because hoods, precoolers, cyclones, fans, motors, and stacks are common to many pollution control systems, they are given extended treatment in the following tentatively numbered separate sections: capture hoods in Section 16, ductwork in Section 7, precoolers (spray chambers and quenchers) in Section 18, cyclones in Section 17, fans and motors in Section 8, and stacks in Section 11. If dust-removal equipment is to be considered, Section 19 discusses screw conveyors.

5.3.2 Total Purchased Cost

The total purchased cost of the fabric filter system is the sum of the costs of the baghouse, bags, auxiliary equipment, instruments and controls; and of taxes and freight. The last three items generally are taken as percentages of the estimated total cost of the first three items. Typical values, from Section 2 of the Manual, are 10% for instruments and controls, 3% for taxes, and 5% for freight.

Bag costs can vary from less than 15% to more than 100% of bare baghouse cost, depending on type of fabric required. This situation makes it inadvisable to estimate total purchased cost without considering both costs, and prevents effective use of factors to estimate a single cost for the baghouse and bags.

5.3.3 Total Capital Investment

Using the Section 2 methodology, the total capital investment (TCI) is estimated from a series of factors applied to the purchased equipment cost to obtain direct and indirect costs for installation. The TCI is the sum of these three costs. The required factors are given in Table 5-7. Because bag costs can have such a large effect on total purchased equipment cost, the factors may cause overestimation of total capital investment when expensive bags are used. Using stainless steel components may also cause overestimation. Because baghouses may vary from small units installed within existing buildings to large, separate structures, specific factors for site preparation or for buildings are not given. However, costs for buildings may be obtained from such references as Means Square Foot Costs 1986.(17) Land, working capital, and offsite facilities are excluded from the table, as they are not normally required. For very large installations, however, they may be needed and would be estimated on an as-needed basis.

Note that the factors given in Table 5-7 are for average installation conditions. Considerable variation may be seen with other-than-average installation circumstances.

5.4 Estimating Total Annual Costs

5.4.1 Direct Annual Cost

Direct annual costs include operating and supervisory labor, operating materials, replacement bags, maintenance (labor and materials), utilities, and dust disposal. Most of these costs are discussed individually below. They vary considerably with location and time, and, for this reason, should be obtained to suit the specific baghouse system being costed. For example, current labor rates may be found in such publications as the Monthly Labor Review, published by the U.S. Department of Labor, Bureau of Labor Statistics.

Table 5-7 Capital Cost Factors for Fabric Filters(8)

Direct Costs		Factor
Purchased equipment costs:		
Fabric filter		As estimated
Bags	A = Sum of	As estimated
Auxiliary equipment		As estimated
Instruments & controls		0.10 A
Taxes		0.03 A
Freight		0.05 A
Total purchased equipment cost	B =	1.18 A
Installation direct costs		
Foundations & supports		0.04 B
Erection & handling		0.50 B
Electrical		0.08 B
Piping		0.01 B
Insulation for ductwork[a]		0.07 B
Painting		0.02 B
Site preparation (S.P.)		As required
Buildings (Bldg.)		As required
Total installation direct costs		0.72 B + S.P. + Bldg.
Total direct costs		1.72 B + S.P. + Bldg.
Indirect costs		
Engineering & supervision		0.10 B
Construction and field expense		0.20 B
Construction fee		0.10 B
Startup fee		0.01 B
Performance test		0.01 B
Contingencies		0.03 B
Total indirect costs		0.45 B
Total direct and indirect costs = Total capital investment		2.17 B + S.P. + Bldg.

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[a] If ductwork dimensions have been established, cost may be estimated based on \$10-12/ft^2 of surface for field application. Fan housings and stacks may also be insulated.[15]

5.4.1.1 Operating and Supervisory Labor

Typical operating labor requirements are 2 to 4 hours per shift for a wide range of filter sizes.(8) Small or well-performing units may require less time, while very large or troublesome units may require more. Supervisory labor is taken as 15% of operating labor.

5.4.1.2 Operating Materials

Operating materials are generally not required for baghouses. An exception is the use of precoat materials injected on the inlet side of the baghouse to provide a protective dust layer on the bags when sticky or corrosive particles might harm them. Adsorbents may be similarly injected when the baghouse is used for simultaneous particle and gas removal. Costs for these materials should be included on a dollars-per-mass basis (e.g., dollars per ton).

5.4.1.3 Maintenance

Maintenance labor varies from 1 to 2 hours per shift.(8) As with operating labor, these values may be reduced or exceeded depending on the size and operating difficulty of a particular unit. Maintenance materials costs are assumed to be equal to maintenance labor costs.(8)

5.4.1.4 Replacement Parts

The major replacement part items are filter bags, which have a normal operating life of 1 to 5 years with about 2 years being typical. The following formula is used for computing the bag replacement cost:

$$CRC_B = (C_B + C_L) \times CRF_B \tag{5-14}$$

where:

CRC_B = bag capital recovery cost (\$/year)

C_B = initial bag cost including taxes and freight (\$)

C_L = bag replacment labor (\$)

CRF_B = capital recovery factor whose value is a function of the annual interest rate and the useful life of the bags. (For instance, for a 10% interest rate and a 2-year life, $CRF_B = 0.5762$.)

The bag replacement labor cost (C_L) will depend on such factors as the number, size, and type of the bags; their accessibility; how they are connected to the baghouse tubesheet; etc. For example, in a reverse-air baghouse it would probably take from 10 to 20 man-minutes to change an 8-in. by 24-ft bag that is clamped in place. This bag has a filtering surface area of approximately 50 ft^2. If the replacement labor rate were \$21.12/h (including overhead), C_L would be from \$0.07 to \$0.14/ft^2 of bag area. As Table 5-6 shows, for some bags (e.g., cotton), this range of C_L would constitute a significant fraction of the purchase cost. For pulse jets, replacement time would be about 5 to 10 man-minutes for a 5-in. by 10-ft bag in a top-access baghouse. These bag replacement times are based on changing a minimum of an entire module and on having typical baghouse designs. Times would be significantly longer if only a few bags were being replaced or if the design for bag attachment or access were atypical.

This method treats the bags as an investment that is amortized over the useful life of the bags, while the rest of the control system is amortized over its useful life (typically 20 years; see Section 5.4.2). Values of CRF_B for bag lives different from 2 years can be calculated from equation (2-3) of the Manual.

5.4.1.5 Electricity

Power is required to operate system fans and cleaning equipment. Fan power for primary gas movement can be calculated from equation (2-7) of the Manual. After substituting into this equation a combined fan-motor efficiency of 0.65 and a specific gravity of 1.000, we obtain:(18)

$$F.P. = 0.000181(Q)(\Delta P)(\theta) \quad (5\text{-}15)$$

where:

F.P. = fan power requirement (kWh/yr)

Q = system flow rate (acfm)

ΔP = system pressure drop (in. H_2O)

θ = operating time (h/yr)

Cleaning energy for reverse-air systems can be calculated from the number of compartments to be cleaned at one time (usually one, sometimes two), and the reverse gas-to-cloth ratio (from about one to two times the forward gas-to-cloth ratio). Reverse-air pressure drop varies up to 6 or 7 in. H_2O depending on location of the fan pickup (before or after the main system fan).(19) The reverse-air fan generally runs continuously.

Typical energy consumption in kWh/yr for a shaker cleaning system operated 8,760 h/yr can be calculated from:(2)

$$P = 0.053A \quad (5\text{-}16)$$

where:

A = gross fabric area (ft^2)

5.4.1.6 Fuel

If the baghouse or associated ductwork is heated to prevent condensation, fuel costs should be calculated as required. These costs can be significant, but may be difficult to predict. For methods of calculating heat transfer requirements, see Perry.(20)

5.4.1.7 Water

Cooling process gases to acceptable temperatures for fabrics being used can be done by dilution with air, evaporation with water, or heat exchange with normal equipment. The last two cases require consumption of plant water, although costs are not usually significant. Section 4.4 of the Manual provides information on estimating cooling-water costs.

5.4.1.8 Compressed Air

Pulse-jet filters use compressed air at pressures of about 60 to 100 psig. Typical consumption is about 2 scfm/1,000 cfm of gas filtered.(2) For example, a unit filtering 20,000 cfm of gas uses about 40 scf of compressed air for each minute the filter is operated.

5.4.1.9 Dust Disposal

If collected dust cannot be recycled or sold, it must be landfilled or disposed of in some other manner. Disposal costs are site-specific, but they may typically run \$20 or \$30 per ton exclusive of transportation (see Section 2.4) of the Manual.

5.4.2 Indirect Annual Cost

These include such costs as capital recovery, property tax, insurance, administrative costs ("G&A"), and overhead. The capital recovery cost is based on the equipment lifetime and the annual interest rate employed. (See Section 2 for a thorough discussion of the capital recovery cost and the variables that determine it.) For fabric filters, the system lifetime varies from 5 to 40 years, with 20 years being typical. However, this does not apply to the bags, which usually have much shorter lives. (See Section 5.4.1.4) Therefore, as Section 2 of the Manual suggests, when figuring the system capital recovery cost, one should base it on the installed capital cost less the cost of replacing the bags (i.e., the purchased cost of the bags plus the cost of labor necessary to replace them). In other words:

$$CRC_s = [TCI - C_B - C_L]\ CRF_s \qquad (5\text{-}17)$$

where:

CRC_s = capital recovery cost for fabric filter system (\$/yr)

TCI = total capital investment (\$)

C_B = initial cost of bags including taxes and freight (\$)

C_L = labor cost for replacing bags (\$)

CRF_s = capital recovery factor for fabric filter system (defined in Section 2).

For example, for a 20-year system life and a 10% annual interest rate, the CRF_s would be 0.1175.

The suggested factor to use for property taxes, insurance, and administrative charges is 4% of the TCI. Finally, the overhead is calculated as 60% of the sum of operating, supervisory, and maintenance labor, and maintenance materials.

5.4.3 Recovery Credits

For processes that can reuse the dust collected in the baghouse or that can sell the dust in a local market, such as fly ash sold as an extender for paving mixes, a credit should be taken. As used below, this credit (RC) appears as a negative cost.

5.4.4 Total Annual Cost

Total annual cost for owning and operating a fabric filter system is the sum of the components listed in Sections 5.4.1 through 5.4.3, i.e.:

$$TAC = DC + IC - RC \qquad (5\text{-}18)$$

where:

TAC = total annual cost (\$)

DC = direct annual cost (\$)

IC = indirect annual cost (\$)

RC = recovery credits (annual) (\$)

5.4.5 Example Problem

Assume a baghouse is required for controlling fly ash emissions from a coal-fired boiler. The flue gas stream is 50,000 acfm at 325 °F and has an ash loading of 4 gr/ft^3. Analysis of the ash shows a mass median diameter of 7 μm. Assume the baghouse operates for 8,640 h/yr (360 d).

Design Gas-to-Cloth Ratio

The gas-to-cloth ratio (G/C) can be taken from Table 5-1 as 2.5, for woven fabrics in shaker or reverse-air baghouses, or 5, for felts used in pulse-jet baghouses. If a factor method were used for estimating G/C, Table 5-3 for shakers would yield the following values: A = 2, B = 0.9, and C = 1.0. The gas-to-cloth ratio would be:

$$2 \times 0.9 \times 1.0 = 1.8.$$

This value could also be used for reverse-air cleaning. For a pulse-jet unit, Table 5-4 gives a value of 9.0 for factor A and 0.8 for factor B. Equation (5-12) becomes:

$$\begin{aligned} V &= 9.0 \times 0.8 \times 2.647\ (275)^{-0.2335} \times (0.7471 + 0.0853 \ln 7) \\ &\quad \times 1.0873\ (4)^{-0.06021} \\ &= 4.69 \end{aligned} \tag{5-19}$$

Because this value is so much greater than the shaker/reverse-air G/C, we conclude that the pulse-jet baghouse would be the least costly design.*

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*This conclusion is based on the inference that a much higher G/C would yield lower capital and, in turn, annual costs. However, to make a more rigorous selection, we would need to calculate and compare the total annual costs of all three baghouse designs (assuming all three are technically acceptable). The reader is invited to make this comparison. Further discussion of the effects of G/C increases, and accompanying pressure drop increases, on overall annual costs will be found in Reference 21.

Assume the use of online cleaning in a common housing structure and, due to the high operating temperature, the use of glass filter bags (see Table 5-5). At a gas-to-cloth ratio of 4.69, the fabric required is:

$$50{,}000 \text{ acfm}/4.69 \text{ fpm} = 10{,}661 \text{ ft}^2.$$

Baghouse Cost

From Figure 5-4, the cost of the baghouse ("common housing" design) is:

$$\text{Cost} = 9{,}688 + 5.552\ (10{,}661) \qquad (5\text{-}20)$$

$$= \$68{,}878$$

Insulation is required. The insulation add-on cost from Figure 5-4 is:

$$\text{Cost} = 1{,}428 + 0.931\ (10{,}661) \qquad (5\text{-}21)$$

$$= \$11{,}353$$

Bag and Cage Cost

From Table 5-6, bag costs are \$1.24/ft^2 for 5-1/8-in diameter glass fiber, bottom removal bags. Total bag cost is:

$$10{,}661 \text{ ft}^2 \times \$1.24/\text{ft}^2 = \$13{,}220.$$

For 10-ft long cages, fabric area per cage = 5-1/8 in./12 in./ft x π x 10 ft = 13.42 ft^2. The number of cages = 10,661 ft^2/13.42 ft^2 = 795 cages. From Table 5-6, individual cage cost is:

$$3.941 + 0.163\ (13.42 \text{ ft}^2) = \$6.128.$$

Total cage cost is:

$$795 \text{ cages} \times \$6.128/\text{cage} = \$4{,}872.$$

Costs of Auxiliaries

Assume the following auxiliary costs have been estimated from data in other parts of the Manual:

Ductwork	$16,000
Fan	16,000
Motor	7,500
Starter	4,000
Dampers	7,200
Screw conveyor	4,000
Stack	8,000
	$62,700

Total Capital Investment

Direct costs for the fabric filter system, based on the factors in Table 5-7, are given in Table 5-8. (Again, we assume site preparation and buildings costs to be negligible.) Total capital investment is $412,315.

Annual Costs--Bags

Table 5-9 gives the direct and indirect annual costs, as calculated from the factors given in Section 5.4. For bag replacement labor, assume 10 min per bag for each of the 795 bags. At a maintenance labor rate of $21.12 (including overhead), the labor cost is $2,809 for 133 h. The bags are assumed to be replaced every 2 yr. The replacement cost is calculated using equation (5-14).

Annual Costs--Pressure Drop

Pressure drop (for energy costs) can be calculated from equations (5-8) through (5-10), with assumed values of 15 [in. $H_2O/(ft/min)]/(lb/ft^2)$ for K_2, 100 psig for P_j, and a cleaning interval of 10 min. We further assume that the G/C (4.69 ft/min) is a good estimate of the mean face velocity over the duration of the filtering cycle.

Table 5-8 Example Costs for Fabric Filter System

Purchased Equipment Costs	
Fabric filter (with insulation)	$ 80,231
Bags and cages	18,092
Auxiliary equipment	62,700
	$161,023 = A
Instruments and controls, 0.1A	16,102
Taxes, 0.03A	4,831
Freight, 0.05A	8,051
Total purchased equipment cost	$190,007 = B
Installation Direct Costs	
Foundation and supports, 0.04B	7,600
Erection and handling, 0.50B	95,004
Electrical, 0.08B	15,201
Piping, 0.01B	1,900
Insulation for ductwork, 0.07B	13,300
Painting, 0.02B	3,800
Site preparation	--
Facilities and buildings	--
Total installation direct costs	$136,805
Total direct costs	$326,812
Indirect Costs	
Engineering and supervision, 0.10B	19,001
Construction and field expense, 0.20B	38,001
Construction fee, 0.10B	19,001
Startup fee, 0.01B	1,900
Performance test, 0.01B	1,900
Contingencies, 0.03B	5,700
Total indirect costs	$85,503
Total capital investment	$412,315

Table 5-9 Example Annual Costs for Fabric Filter Systems

Direct Annual Costs	
Operating labor	
Operator, 6 h/day x 360 d/yr x $12/h =	$25,920
Supervisor, 15% of operator =	3,888
Operating materials	--
Maintenance	
Labor, 3 h/day x 360 d/yr x $13.20/h =	14,256
Material, equal to labor costs	14,256
Replacement parts, bags, [2,809 + (13,220 x 1.08*)] x 0.5762 =	9,845
Utilities	
Electricity, 0.000181 x 50,000 acfm x 10.3 in. H_2O x 8,640 h/yr x $0.06/kWh =	48,323
Compressed air (dried and filtered), 2 scfm/1,000 acfm x 50,000 acfm x $0.16/1,000 scfm x 60 min/h x 8,640 h/yr =	8,294
Waste disposal, at $20/ton onsite for essentially 100% collection efficiency:	
$4\ \frac{\text{gr}}{\text{ft}^3}\ \frac{1\ \text{lb}}{7{,}000\ \text{gr}} \times \frac{50{,}000\ \text{ft}^3}{\text{min}} \times \frac{60\ \text{min}}{\text{h}} \times \frac{8{,}640\ \text{h}}{\text{yr}} \times \frac{1\ \text{ton}}{2{,}000\ \text{lb}} \times \frac{\$20}{\text{ton}} =$	148,114
Total direct annual costs	$272,896
Indirect Annual Costs	
Overhead, 0.6 x (25,920 + 3,888 + 14,256 + 14,256) =	34,992
Property tax, 0.01 x 412,315 =	4,123
Insurance, 0.01 x 412,315=	4,123
Administration, 0.02 x 412,315=	8,246
Capital recovery cost, (412,315 - 2,809 - 13,220 x 1.08) x 0.1175	46,439
Total indirect annual costs	97,923
Total annual cost	$371,000 (rounded)

*The "1.08" accounts for freight and sales tax on the bags.

$$W_o = C_i V\theta = 4\ \text{gr/ft}^3 \times \frac{1\ \text{lb}}{7{,}000\ \text{gr}} \times 4.69\ \text{ft/min} \times 10\ \text{min} = 0.0268\ \text{lb/ft}^2 \qquad (5\text{-}22)$$

$$\Delta P = 6.08 \times 4.69\ \text{ft/min} \times [100\ \text{psig}]^{-0.65} + 15\ [\text{in. } H_2O/\text{ft/min})]/(\text{lb/ft}^2)$$

$$\times\ 0.0268\ \text{lb/ft}^2 \times 4.69\ \text{ft/min} = 3.32\ \text{in. } H_2O$$ across the fabric (when fully loaded).

Assume the baghouse structure and the ductwork contribute an additional 3 in. H_2O and 4 in. H_2O, respectively. The total pressure drop is, therefore, 10.3 in.

Total Annual Cost

The total annual cost is $371,000, nearly half of which is for ash disposal. If a market for the flyash could be found, the total annual cost would be greatly reduced. For example, if $2/ton were received for the ash, the total annual cost would drop to $208,000 ($370,819 - $148,114 - $14,811), or 56% of the cost when no market exists. Clearly, the total annual cost is extremely sensitve to the value chosen for the dust disposal cost in this case. In this and in similar cases, this value should be selected with care.

5.4.6 Acknowledgment

We gratefully acknowledge the following companies for contributing data to this section:

Aget

BACT

BHA

Dustex

Fuller

Griffin

W. W. Sly

Zurn

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